

PB-242 572

PROCEEDINGS: SYMPOSIUM ON FLUE GAS DESULFURIZATION,
HELD IN ATLANTA, GEORGIA, ON NOVEMBER 1974. VOLUME I

NATIONAL ENVIRONMENTAL RESEARCH CENTER

DECEMBER 1974

DISTRIBUTED BY:

NTIS

National Technical Information Ser.
U. S. DEPARTMENT OF COMMERCE

149056
EPA-650/2-74-126-a

PB 242 572

DECEMBER 1974

Environmental Protection Technology Series

**PROCEEDINGS:
SYMPOSIUM
ON FLUE GAS DESULFURIZATION -
ATLANTA, NOVEMBER 1974
VOLUME I**



Reproduced by
**NATIONAL TECHNICAL
INFORMATION SERVICE**
U.S. Department of Commerce
Springfield, VA. 22151

Office of Research and Development
U.S. Environmental Protection Agency
Washington, DC 20460

PB242572

TECHNICAL REPORT DATA (Please read instructions on the reverse before completing)		
1. REPORT NO. EPA-650/2-74-126-a	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Proceedings: Symposium on Flue Gas Desulfurization-- Atlanta, November 1974		5. REPORT DATE December 1974
7. AUTHOR(S) Miscellaneous		6. PERFORMING ORGANIZATION CODE
9. PERFORMING ORGANIZATION NAME AND ADDRESS NA		8. PERFORMING ORGANIZATION REPORT NO.
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development NERC-RTP, Control Systems Laboratory Research Triangle Park, NC 27711		10. PROGRAM ELEMENT NO. LAB013: ROAP 2IACX-AA
		11. CONTRACT/GRANT NO. In-House
		13. TYPE OF REPORT AND PERIOD COVERED Proceedings: 11/4-7/74
		14. SPONSORING AGENCY CODE
15. SUPPLEMENTARY NOTES		
16. ABSTRACT The proceedings document the presentations made during the symposium, which dealt with the status of flue gas desulfurization technology, both in the U.S. and abroad. The presentations emphasize process costs, both regenerable and non-regenerable processes, second generation processes, and byproduct disposal/utilization. Aim of the symposium was to provide potential users of sulfur oxide control technology with a current review of progress made in applying processes for the reduction of sulfur oxide emissions at the full- or semi-commercial scale. The symposium was the sixth of such EPA-sponsored meetings, dating back to 1966, when the topic of principal concern was the use of limestone to control sulfur oxide emissions.		
17. KEY WORDS AND DOCUMENT ANALYSIS		
a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution Flue Gases Desulfurization Sulfur Oxides Cost Effectiveness Regeneration (Engineering)	Byproducts Disposal Marketing Sludge	Air Pollution Control Stationary Sources
		13B 21B 07A, 07D, 05C 07B 14A
18. DISTRIBUTION STATEMENT Unlimited		19. SECURITY CLASS (This Report) Unclassified
		20. SECURITY CLASS (This page) Unclassified
		21. NO. OF PAGES 661
		22. PRICE 15.25

ATTENTION

AS NOTED IN THE NTIS ANNOUNCEMENT,
PORTIONS OF THIS REPORT ARE NOT LEGIBLE.
HOWEVER, IT IS THE BEST REPRODUCTION
AVAILABLE FROM THE COPY SENT TO NTIS.

PB242572

EPA-650/2-74-126-a

**PROCEEDINGS:
SYMPOSIUM
ON FLUE GAS DESULFURIZATION -
ATLANTA, NOVEMBER 1974
VOLUME I**

ROAP NO. 21ACX-AA
Program Element No. 1AB013

Chairman: E.L. Plyler
Vice-Chairman: W.H. Ponder

Sponsored by
Control Systems Laboratory
National Environmental Research Center
Research Triangle Park, N. C. 27711

Prepared for
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C. 20460

December 1974

ia

EPA REVIEW NOTICE

This report has been reviewed by the National Environmental Research Center - Research Triangle Park, Office of Research and Development, EPA, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into series. These broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and maximum interface in related fields. These series are:

1. ENVIRONMENTAL HEALTH EFFECTS RESEARCH
2. ENVIRONMENTAL PROTECTION TECHNOLOGY
3. ECOLOGICAL RESEARCH
4. ENVIRONMENTAL MONITORING
5. SOCIOECONOMIC ENVIRONMENTAL STUDIES
6. SCIENTIFIC AND TECHNICAL ASSESSMENT REPORTS
9. MISCELLANEOUS

This report has been assigned to the ENVIRONMENTAL PROTECTION TECHNOLOGY series. This series describes research performed to develop and demonstrate instrumentation, equipment and methodology to repair or prevent environmental degradation from point and non-point sources of pollution. This work provides the new or improved technology required for the control and treatment of pollution sources to meet environmental quality standards.

This document is available to the public for sale through the National Technical Information Service, Springfield, Virginia 22161.

Publication No. EPA-650/2-74-126-a

PREFACE

The development of technologies to control sulfur dioxide emissions is an issue of national importance. Indications that the use of sulfur-containing fossil fuels to generate electricity will increase by 50 percent by 1985, recognition that more than half of all sulfur dioxide (SO₂) emissions are caused by electric power generation, and documentation of the deleterious effects caused by such emissions make the acceleration of development to commercial application of SO₂ control technology one of the most important goals of the U.S. Environmental Protection Agency (EPA). The Control Systems Laboratory (CSL), as part of EPA's Office of Research and Development, has accelerated the development of flue gas desulfurization (FGD) technology so that it is now in the process of commercialization. Each year, CSL sponsors a Symposium which brings together users and developers of the technology for open exchange and discussion of their experience and results. The 1974 Symposium was held on November 4-7 at the Sheraton-Biltmore Hotel in Atlanta.

Commercial application of full-scale FGD processes was the primary topic of the Symposium which provided a means to transfer the latest information on the status, development and assessment of current FGD processes to developers, vendors, potential users, and those concerned with regulatory deadlines and enforcement. The Symposium was attended by more than 600 national and international representatives of utilities, vendors, government, and universities. One-third of the presentation was assessment and status reports from utilities which are utilizing full-scale FGD systems. Information from small- and large-scale pilot testing of alternate FGD systems, developments in FGD by-product disposal and utilization, and cost studies of FGD application completed the extensive overview of FGD technology.

In concluding remarks, it was noted by an international FGD technology consultant that the Symposium is recognized as the world's most significant conference on SO₂ pollution control by FGD technology. The growth of the FGD Symposium to this latest, widely attended, in-depth information exchange forum supports the conclusion that FGD is the only viable near-term alternative, other than the burning of clean fuels, that will permit compliance with current regulatory requirements.

The contents of these Proceedings are comprised of copies of the participating authors' papers as received. Although the papers have been reviewed and approved for publication by the Environmental Protection Agency, approval does not indicate that the contents necessarily reflect the views and/or policies of the Agency. The mention of trade names or commercial products does not constitute endorsement or recommendation for use.

As supplies permit, copies of the Proceedings are available free of charge and may be obtained by contacting the Air Pollution Technical Information Center, Environmental Protection Agency, Research Triangle Park, North Carolina 27711.

CONTENTS

<u>TITLE</u>	<u>PAGE</u>
VOLUME I	
OPENING SESSION	
Keynote Address - THE ROLE OF ENVIRONMENTAL HEALTH ASSESSMENT IN THE CONTROL OF AIR POLLUTION John Finklea, Environmental Protection Agency, National Environmental Research Center, Research Triangle Park, North Carolina	1
FLUE GAS DESULFURIZATION AND OTHER ALTERNATIVES FOR PRODUCING ELECTRICITY FROM COAL S.J. Gage, Environmental Protection Agency, Washington, D. C.	5
STATUS OF FLUE GAS DESULFURIZATION SYSTEMS IN THE UNITED STATES T.W. Devitt and F. Zada, PEDCo-Environmental, Inc., Cincinnati, Ohio	17
STATUS OF FLUE GAS DESULFURIZATION TECHNOLOGY IN JAPAN J. Ando, Chuo University, Tokyo, Japan	125
COST COMPARISONS OF FLUE GAS DESULFURIZATION SYSTEMS G.G. McGlamery and R.L. Forstrick, Tennessee Valley Authority, Muscle Shoals, Alabama	149
NON-REGENERABLE PROCESSES SESSION	
EPA/RTP PILOT STUDIES RELATED TO UNSATURATED OPERATION OF LIME AND LIMESTONE SCRUBBERS R.H. Borgwardt, Environmental Protection Agency, Research Triangle Park, North Carolina	225
LIMESTONE AND LIME TEST RESULTS AT THE EPA ALKALI SCRUBBING TEST FACILITY AT THE TVA SHAWNEE POWER PLANT M. Epstein, Bechtel Corporation, San Francisco, California	241
OPERATIONAL STATUS AND PERFORMANCE OF THE ARIZONA PUBLIC SERVICE COMPANY FLUE GAS DESULFURIZATION SYSTEM AT THE CHOLLA STATION L.K. Mundth, Arizona Public Service, Phoenix, Arizona	307
WET SCRUBBER OPERATING EXPERIENCE AT LA CYGNE STATION UNIT NO. 1 C.F. McDaniel, Kansas City Power and Light, Kansas City, Missouri	319

Preceding page blank

<u>TITLE</u>	<u>PAGE</u>
DUQUESNE LIGHT COMPANY, PHILLIPS POWER STATION, LIME SCRUBBING FACILITY S.L. Pernick, Jr. and R.G. Knight, Duquesne Light Company, Pittsburgh, Pennsylvania	329
THE HORIZONTAL CROSS FLOW SCRUBBER A. Weir, Jr., J.M. Johnson, D.G. Jones, and S.T. Carlisle, Southern California Edison, Rosemead, California	357
OPERATIONAL STATUS AND PERFORMANCE OF THE LOUISVILLE FGD SYSTEM AT THE PADDY'S RUN STATION R.P. Van Ness, Louisville Gas and Electric Company, Louisville, Kentucky	389
DISPOSAL OF BY-PRODUCTS FROM NON-REGENERABLE FLUE GAS DESULFURIZATION SYSTEMS J. Rossoff, R.C. Rossi, L.J. Bornstein, The Aerospace Corporation, El Segundo, California J.W. Jones, Environmental Protection Agency, Research Triangle Park, North Carolina	399
AN OVERVIEW OF DOUBLE ALKALI PROCESSES FOR FLUE GAS DESULFURIZATION N. Kaplan, Environmental Protection Agency, Research Triangle Park, North Carolina	445
INITIAL OPERATING EXPERIENCES WITH A DUAL-ALKALI SO ₂ REMOVAL SYSTEM PART I - PROCESS PERFORMANCE WITH A COMMERCIAL DUAL-ALKALI SO ₂ REMOVAL SYSTEM T.T. Dingo, General Motors Corporation, Parma, Ohio	517
PART II - EQUIPMENT PERFORMANCE WITH A COMMERCIAL DUAL-ALKALI SO ₂ REMOVAL SYSTEM E.J. Piasecki, General Motors Corporation, Parma, Ohio	539
EPA-ADL DUAL ALKALI PROGRAM--INTERIM RESULTS C.R. LaMantia, R.R. Lunt, J.E. Oberholtzer, E.L. Field, Arthur D. Little, Inc., Cambridge, Massachusetts N. Kaplan, Environmental Protection Agency, Research Triangle Park, North Carolina	549

TITLEPAGE

VOLUME II

REGENERABLE PROCESSES SESSION

NEW ENGLAND SO ₂ CONTROL PROJECT FINAL RESULTS G.R. Koehler, E.J. Dober, Chemical Construction Corporation, New York, New York	671
ASSESSMENT OF PROTOTYPE OPERATION AND FUTURE EXPANSION STUDY - MAGNESIA SCRUBBING MYSTIC GENERATING STATION C.P. Quigley, J.A. Burns, Boston Edison Company, Boston, Massachusetts	709
MAG-OX SCRUBBING EXPERIENCE AT THE COAL-FIRED DICKERSON STATION, POTOMAC ELECTRIC POWER COMPANY D.A. Erdman, Potomac Electric Power Company, Washington, D.C.	729
POWER PLANT FLUE GAS DESULFURIZATION BY THE WELLMAN-LORD SO ₂ PROCESS PART I - THE DEAN H. MITCHELL STATION E.L. Mann, Northern Indiana Public Service Company, Michigan City, Indiana	739
PART II - CONTINUING PROGRESS FOR THE WELLMAN-LORD SO ₂ PROCESS E.E. Bailey, Davy Powergas, Inc., Lakeland, Florida	745
THE CAT-OX DEMONSTRATION PROGRAM E.M. Jamgochian, The Mitre Corporation, McLean, Virginia W.E. Miller, Illinois Power Company, Decatur, Illinois	761
THE SHELL FLUE GAS DESULFURIZATION PROCESS J.B. Pohlenz, Universal Oil Products Company, Des Plaines, Illinois	807
STATUS REPORT ON CHIYODA THOROUGHbred 101 PROCESS M. Noguchi, Chiyoda International Corporation, Seattle, Washington	837
FLUE GAS DESULFURIZATION BY-PRODUCT DISPOSAL/UTILIZATION PANEL	
FLUE GAS DESULFURIZATION BY-PRODUCT DISPOSAL/UTILIZATION - REVIEW AND STATUS H.W. Elder, Tennessee Valley Authority, Muscle Shoals, Alabama	851

<u>TITLE</u>	<u>PAGE</u>
LIME/LIMESTONE SLUDGE DISPOSAL - TRENDS IN THE UTILITY INDUSTRY C.N. Ifendi and H.S. Rosenberg, Battelle, Columbus, Ohio	865
ENVIRONMENTALLY ACCEPTABLE DISPOSAL OF FLUE GAS DESULFURIZATION SLUDGES: THE EPA RESEARCH AND DEVELOPMENT PROGRAM J.W. Jones, Environmental Protection Agency, Research Triangle Park, North Carolina	887
FGD SLUDGE FIXATION AND DISPOSAL W.H. Lord, Dravo Corporation, Pittsburgh, Pennsylvania	929
UTILIZING AND DISPOSING OF SULFUR PRODUCTS FROM THE GAS DESULFURIZATION PROCESSES IN JAPAN J. Ando, Chuo University, Tokyo, Japan	955
TVA-EPA STUDY OF THE MARKETABILITY OF ABATEMENT SULFUR PRODUCTS J.I. Bucy and P.A. Corrigan, Tennessee Valley Authority, Muscle Shoals, Alabama	969
THE PRODUCTION AND MARKETING OF SULFURIC ACID FROM THE MAGNESIUM OXIDE FLUE GAS DESULFURIZATION PROCESS I.S. Zonis, F. Olmsted, K.A. Holst, D.M. Cunningham, Essex Chemical Corporation, Clifton, New Jersey	1003
SECOND GENERATION PROCESSES SESSION	
SECOND GENERATION PROCESSES FOR FLUE GAS DESULFURIZATION - INTRODUCTION AND OVERVIEW A.V. Slack, SAS Corporation, Sheffield, Alabama	1029
PILOT PLANT TESTING OF THE CITRATE PROCESS FOR SO ₂ EMISSION CONTROL W.A. McKinney, W.I. Nissen, D.A. Elkins and J.B. Rosenbaum, Bureau of Mines, Salt Lake City, Utah	1049
TVA-EPA PILOT-PLANT STUDY OF THE AMMONIA ABSORPTION - AMMONIUM BISULFATE REGENERATION PROCESS C.E. Breed, Tennessee Valley Authority, Muscle Shoals, Alabama G.A. Hollinden, Tennessee Valley Authority, Chattanooga, Tennessee	1069

TITLE

PAGE

DESCRIPTION AND OPERATION OF THE STONE & WEBSTER/IONICS
SO₂ REMOVAL AND RECOVERY PILOT PLANT AT THE WISCONSIN
ELECTRIC POWER COMPANY VALLEY STATION IN MILWAUKEE

K.A. Meliere and R.J. Cartside, Stone & Webster
Engineering Corporation, Boston, Massachusetts

W.A. McRae and T.F. Seamans, Ionics, Inc.,
Waltham, Massachusetts

1109

CALSOX SYSTEM DEVELOPMENT PROGRAM PRESENTED AT
THE EPA FLUE GAS DESULFURIZATION SYMPOSIUM

R.E. Barnard and R.K. Teague, Monsanto
Enviro-Chem Systems, Inc., St. Louis, Missouri

G.C. Vansickle, Indianapolis Power & Light Company,
Indianapolis, Indiana

1127

WESTVACO ACTIVATED CARBON PROCESS FOR SO_x
RECOVERY AS ELEMENTAL SULFUR

F.J. Ball, G.N. Brown, A.J. Repik, S.L. Torrence,
Westvaco Corporation, North Charleston,
South Carolina

1151

THE ROLE OF ENVIRONMENTAL HEALTH ASSESSMENT
IN THE CONTROL OF AIR POLLUTION

Dr. John Finklea
Director
National Environmental Research Center
Research Triangle Park
Environmental Protection Agency

Presented at
Symposium on Flue Gas Desulfurization
Sponsored by
Control Systems Laboratory
U.S. Environmental Protection Agency
Atlanta, Georgia
November 4-7, 1974

THE ROLE OF ENVIRONMENTAL HEALTH ASSESSMENT
IN THE CONTROL OF AIR POLLUTION

ABSTRACT*

Existing legislation requires that air pollution be controlled to protect public health. The health assessments required to evaluate air quality standards and to support decisions on the control of mobile and stationary source emissions present a series of complex scientific challenges.

Assessment of the health-related air quality standards reveals no reason for abruptly changing these air quality goals. However, vexing uncertainties in the scientific information base remain. Thus there are often rather large differences between upper and lower boundary estimates of a threshold for adverse health effects. In addition, there is a substantial body of evidence that fine particulate aerosols like the acid-sulfate aerosols are injurious, but the broad scientific information base necessary to establish aerosol standards is lacking.

Assessment of the health effects of increasing sulfur oxide emissions is likewise difficult because of the uncertainties involved in our scientific information base. However, present rough estimates conclude that substantial excess adverse health effects can be expected each year if clean air requirements are not met: thousands of premature deaths, millions of days of illness among susceptible segments of the population, hundreds of thousands of needless acute lower respiratory illnesses in otherwise healthy children and hundreds of thousands of chronic respiratory disorders among adults. If the health impact of short-term fumigations prove greater than expected or if regional effects occur in the western power regions, the calculations of excess adverse effects given here may prove overly conservative. It is important to remember that the present ambient sulfur dioxide standards would be about what is necessary to protect public health from effects attributable to acid sulfate aerosols if dispersion conditions were good and if acid-aerosols did not intrude from up wind sources. One should also recall that the sulfur oxides criteria document recognized the problem of acid aerosols but that the knowledge at that time led one to believe that long range transport of aerosols was not a major constraint for air quality.

*Full text to be published in Advances in Environmental Science & Technology, Volume 7, published by Wiley-Interscience, John Wiley & Sons, Inc., New York, after July 1975.

Assessment of the health impact of equipping light duty motor vehicles with oxidation catalysts demonstrates that these emission control devices are at best a mixed blessing. Emissions of a number of pollutants capable of adversely effecting public health including carbon monoxide, phenols, aldehydes and polynuclear aromatics are dramatically reduced by these devices. Likewise, catalytic converters will reduce other hydrocarbon emissions which are not known to affect health adversely but which are an important precursor of photochemical oxidants that do adversely affect health. Unfortunately, oxidation catalysts will also alter emissions patterns so that sulfates and sulfuric acid emissions will be greatly increased and worsen a public health problem whose dimensions are not completely understood. The degree of public controversy engendered by the health assessments just described is a function of the scientific uncertainties contained in each particular assessment. Another major determinant of the amount of controversy involved is whether or not the decision affects a vested industrial, environmental or governmental interest that is politically and economically potent. Since control costs usually rise exponentially as one approaches an emission or ambient standard, uncertainties about exposure response relationships can result in violent controversy and cause major economic problems. In general, the amount of scientific information demanded seems directly related to the degree of public controversy. Despite attempts to augment the scientific information base for air pollution control, it is unlikely that scientific uncertainties will be sufficiently resolved to prevent disruptive disputes among reasonable persons for another decade.

FLUE GAS DESULFURIZATION AND OTHER ALTERNATIVES
FOR PRODUCING ELECTRICITY FROM COAL

Dr. Stephen Gage
Acting Director
Office of Energy Research
Environmental Protection Agency

Presented at
Symposium on Flue Gas Desulfurization
Sponsored by
Control Systems Laboratory
U.S. Environmental Protection Agency
Atlanta, Georgia
November 4 - 7, 1974

Preceding page blank

Since the last symposium on flue gas cleaning, there have been a number of developments of immense proportions in the energy area. Some of these developments have significant implications for this meeting and, more specifically, for the subject of my presentation.

In June 1973, the President committed the nation to a much expanded Federal research and development program on energy supply and use. An interagency study chaired by the Atomic Energy Commission recommended in December 1973 that \$10 billion dollars be spent over the next five years on energy R&D. The AEC's report, "The Nation's Energy Future," placed greatest emphasis on coal and nuclear power. The general thrust of the report was adopted by the Office of Management and Budget in formulating the FY 1975 energy R&D budget which nearly doubled the FY 1974 level.

To bring improved efficiency to the Federal energy R&D effort, the Administration proposed in July 1973 that an Energy Research and Development Administration be formed to provide centralized management of nuclear, fossil fuel, and other R&D. The ERDA act was passed in October 1974 and soon the Atomic Energy Commission and some programs of the Department of Interior and National Science Foundation will be merged in the new organization.

October 1973 brought the Arab oil embargo which was followed by the gasoline crisis, Project Independence, and the Federal Energy Administration. The long queues of cars at gasoline stations and the higher prices posted on the pumps brought home the realization of how vulnerable we were becoming to international political blackmail. Project Independence committed us on a course to reduce that vulnerability by increasing our capability for energy self-sufficiency. That course, as the "Project Independence Blueprint" will tell us soon, is a complex one requiring many different types of actions -- demand reduction, increased use of coal, offshore exploration for oil and gas, etc.

Perspective on Energy Self-Sufficiency

Research and development on new energy resources and new energy conversion technologies will unfortunately have little impact for a good many years on our capability for energy self-sufficiency. The only significant advances toward this national goal will come from application of technologies and practices which either exist now or which have been under development for a number of years.

Near the top of any reasoned agenda for energy policy actions is energy conservation. We are making energy conservation work today and, of course, are being helped considerably by the significant increases in the prices of all forms of energy during the past year. Not only oil but coal, natural gas and electricity have increased sharply in price. We are using energy less profligately and also more efficiently. Smaller cars with fewer power-consuming options, less wasteful use of energy in heating and cooling of homes and offices, more efficient use of energy in industrial processes -- all confirm that energy conservation is an alternative and a practical and necessary one at that.

On the supply side of the equation, we can anticipate increased use of nuclear power, domestic oil and gas, coal, shale oil, geothermal, and solar energy. Nuclear power expansion through the middle 1980's, however, is tied to established construction schedules; there will be at most slight accelerations or delays from those schedules. Rapid development of additional oil and gas supplies is probably limited to expanded drilling near existing fields and gathering systems and to advanced recovery techniques applied to existing wells. There is undoubtedly much more oil on the Alaskan North Slope and in new offshore areas but exploration for and development of those resources will require years. Shale oil, geothermal and solar will provide some new supplies but within the coming decade their contribution will be marginal at best. That leaves one major near-term alternative to imported oil: combustion of coal in utility and industrial boilers -- boilers which are either operating or are already under construction.

The serious implications of expanded coal consumption for the nation's air quality brings us back to the subject of this symposium. In my opinion, there are no alternatives to the use of both high and low sulfur coal. Both must be used to meet the serious and continuing fuel supply crisis which faces the nation. But both must be used in ways which do not seriously jeopardize human health and welfare. The question then is how we use the nation's different types of coal in environmentally acceptable ways.

Although generally of higher heating quality, coal from the eastern and central part of the nation typically has higher sulfur content. Coal from the Interior Province of Western Kentucky, Illinois and Indiana often has a sulfur content of three percent or greater, thus requiring sulfur oxides control greater than 80 percent to meet New Source Performance Standards and many state regulations. Appalachian coal varies considerably from metallurgical

quality coal with sulfur content well below one percent to medium quality coal with sulfur content between two and three percent. Recent estimates of low sulfur Appalachian coal reserves have, however indicated much larger reserves than previously thought. The Geological Survey has estimated that there are about 100 billion tons of coal with sulfur content less than one percent in the Appalachian fields, primarily in southern West Virginia, eastern Kentucky and western Virginia. Much of this coal will meet NSPS and if the rest is washed it too can meet the most stringent emission standards in middle central states.

Because of the increased price of coal following the oil embargo, I am confident that increasing amounts of this high quality eastern coal will be produced for use in both utility and industrial boilers. Today's differential between production costs and selling price of coal is sufficient to provide for full reclamation of strip mined areas and adequate health and safety devices in underground mines. Thus much of the boiler fuel required in the industrial heartland will be low sulfur coal from the prime Appalachian fields with acceptable levels of impacts to the air and land of the region.

Poorer quality Appalachian and nearly all Interior coal will require sulfur control. This fact, of course, underlies the *raison d'être* of both the Environmental Protection Agency's flue gas desulfurization R&D program and the Department of Interior's (soon to be ERDA's) synthetic fuel R&D program.

Further justification for the massive Federal R&D effort on synthetic fuels arises from the limited degree to which coal can be substituted for other fuels. To make the point abundantly clear, use of the vast reserves of Western coal depends to a considerable extent on the development of technology to convert that low sulfur Western coal to alternative fuel forms -- clean liquids and gases -- and, not incidentally, on the development of technology to remove the ash and sulfur in the process.

Western coals, typically low in both sulfur and heat content, are not, I'm afraid, the panacea that some believe. In the next five years, Western coals will provide important new supplies but they are not a widely applicable solution to coal supply and use east of the Mississippi River where most of America's heavy industry is located. Transportation of the coal from the Wyoming and Montana coal fields to Chicago and points east is expensive and consumes valuable and scarce diesel fuel. Adding \$10 to \$15 or more per ton to the price of coal for transportation will in many cases tip the balance toward eastern coals with sulfur control measures.

There are other problems with using Western coals in existing Midwestern boilers. Use of western coal in many of these boilers would result in the derating of the boilers by 10 to 15 percent because of limits on bulk handling capabilities of the entire plant. With up to 30 percent lower heat content and higher ash and moisture content, western coals would cause a much more serious reduction of reserve generating margin than stack gas scrubbers. There are operational problems with western coals, such as preheater fires due to carbon carryover and overloading of particulate collection devices. Finally, a significant fraction of the western coals will not meet the more stringent emission standards and in some cases the New Source Performance Standards.

Western coals, I believe, will be important but will only begin making significant contributions when they can be used as feedstock for coal gasification and liquefaction process. Now let me turn to those processes.

Alternative Technologies To Stack Gas Scrubbing

In the paper which I delivered before the last Flue Gas Desulfurization Symposium (May 1973), I discussed in detail the precombustion cleaning processes which might be considered as alternatives to flue gas desulfurization. I also mentioned the potential of fluidized bed combustion as an alternative but, because of the low level of R&D activity in this area at that time, I did not expand on fluidized bed systems.

The political and economic climate influencing the development of alternative technologies has changed drastically in the past year and a half. In the FY 1975 budget, \$350 million were made available to the Office of Coal Research and Bureau of Mines for coal utilization R&D. The two technologies which appear to be the most direct competitors of flue gas desulfurization -- low-Btu gasification and fluidized bed combustion -- received the greatest proportional increases over FY 1974. Low-Btu gasification R&D received \$52 million and fluidized bed combustion R&D received \$37 million in the FY 1975 budget. These funding levels are comparable to the \$35 million allocated to the Environmental Protection Agency for stack gas cleaning R&D in FY 1975.

Research and development, however, must proceed apace. Adequate resources are a prerequisite for orderly progress in an R&D program, however, progress depends on many other factors. The advance of a project through the R&D stream cannot out-run ideas, manpower, and construction schedules. Thus, while it appears likely that low-Btu gasification and fluidized bed combustion will receive adequate support, actual as opposed to paper progress in those alternatives has been slight. Further, there have been few technical developments in the R&D on liquefaction and high-Btu gasification not fully anticipated a year and a half ago, even though those two processes have also received marked funding increases.

Before discussing low-Btu gasification and fluidized bed combustion further, let me put liquefaction and high-Btu gasification in perspective relative to electric power generation. There appears to be an emerging consensus that both liquefaction and high-Btu gasification will produce fuels that will be too expensive for use under utility boilers. Production costs for liquid coal products range from \$1.10 to \$2.50 per million Btu while costs for high-Btu synthetic gas range from \$1.20 to \$1.80 per million Btu. When these synthetic fuels become available in any quantity -- and that will be in the 1980's -- I believe that they will be bid away from the utility market by industrial, commercial, and residential markets where clean liquid and gaseous fuels will seek premium uses.

From the point of view of resource conservation, it also makes more sense to utilize the clean fuels derived from these two processes in high efficiency end uses such as residential heating. Coal liquefaction, it is estimated, will be 60 to 70 percent thermally efficient while high-Btu gasification will be 55 to 65 percent efficient. It would be illogical to waste 60 percent or more of the remaining heat value in the cleaned fuel by converting it to electricity.

Of the various clean fuel alternatives, these two are the most advanced. The hydrogasification pilot plant being tested by the Institute for Gas Technology in Chicago has had several long successful runs in the past few months. The carbon dioxide acceptor pilot plant operated by Consolidated Coal Company in Rapid City, South Dakota has also had numerous successful runs, exceeding the design specifications of the gasifier. The synthane and BIGAS pilot plants are under construction in Bruceton and Homer City, Pennsylvania, respectively. The most advanced coal liquefaction process -- Pittsburgh and Midway's Solvent Refined Coal process -- will soon be tested in a pilot plant near Fort Lewis, Washington.

R&D efforts for both low-Btu gasification and fluidized bed combustion lag those for high-Btu gasification and liquefaction. The development of a 5-year fossil fuel R&D program by the Interior Department's Office of Research and Development during summer 1974 helped to clarify some of the goals and program thrusts for these two technologies but many questions remain. The program plan calls for a low-Btu gasification demonstration plant in 1978 and fluidized bed combustion demonstration in 1979 -- goals which I submit are probably impossible to meet, given the state of the technology and the managerial and technical expertise supporting the efforts.

One of the toughest questions to be answered in both R&D programs is which way(s) to go. With the realization that sulfur control measures will be required for electrical power plants and that these may be alternatives to stack gas cleaning, there have been a plethora of proposals on candidate processes. So much so that in the area of low-Btu gasification the number of candidates is overwhelming and the adequacy of the screening process -- that is, selection criteria based on technical knowledge -- is frankly underwhelming. Hopefully, these organizational problems will soon be overcome so that both programs can move ahead. One good sign is the development of a National Fluidized Bed Combustion R&D Plan designed to define the program goals, translate these into research plans, and spell out the roles of the various agencies. EPA is participating in the development of the plan because of the agency's important contributions to the early development of fluidized bed boilers and because of the agency's continuing role in the environmental assessment of these combustors.

The advantages of low-Btu gasification and fluidized bed combustion have been widely touted in the past year. Both techniques have the potential for improving the thermal efficiency of electrical power production. Used in a combined cycle configuration, i.e., with a gas turbine topping cycle, the overall plant efficiency may go as high as 45 percent. But such high efficiencies will be achievable only in second or third generation plants. Early low-Btu gasification efficiency and early fluidized bed boilers will probably not achieve over 36 to 37 percent. And those efficiencies assume success in the single most critical R&D element -- hot gas cleanup. Thus it is likely that early low-Btu gasification and fluidized bed combustion plants will use coal with approximately the same efficiency as a modern steam boiler with a low pressure drop stack gas scrubber.

Another obvious point which must be made is that since the low-Btu gas plant, the fluidized bed boiler, and the conventional boiler with stack gas scrubber are all located at the central electrical generating station they are all faced with the common problem of sulfur. Assuming that technologies will develop to the point where the sulfur can be removed from the coal, the nagging problem of sulfur by-product disposal remains. If dolomite, limestone, or lime is used to remove the sulfur, then landfill techniques or expensive and unproven regeneration techniques must be used to handle the by-products. If, from the reducing environment of the gasifier, the sulfur can be removed as hydrogen sulfide it can then be converted to elemental sulfur in a Claus process. Advanced flue gas desulfurization processes, to which I'll return later, also have the potential for producing elemental sulfur. The problem of disposing of the elemental sulfur may not be trivial, however.

Predictions of low production costs for both low-Btu gasification and fluidized bed combustion have also been touted by their proponents as justification for rapid development and deployment. Estimates of low-Btu power gas production costs have gone up in the past year, reflecting more the rising estimates for high-Btu gasifiers than detailed economic analyses of low-Btu gasifier designs. Estimates of incremental costs for power gas production range from 50 to 80 cents per million Btu based on a 90% plant factor. Although proponents of fluidized bed combustion claim that the process may in fact be less expensive than conventional pulverized coal-burning combustion, such cost savings could not be approached for the first several generations of combustors, that is, until late in the 1980's. One component of FBC costs which I believe has been seriously underestimated is the waste disposal or regeneration of the reacted dolomite. Those of you who have been working on stack-gas scrubber waste disposal should certainly appreciate that fact.

Finally, I should make a categorical statement about the trend of estimated costs of almost any new technology as the technology moves from the bench scale to the pilot plant to the demonstration plant -- the estimated costs invariably go up and they go up more rapidly as the demonstration phase is approached. There is nothing mysterious about this -- it's human nature. In the early days of nuclear power the wiser and experienced engineers always multiplied the cost estimates of new reactor concepts by a factor of three. The more mathematically oriented French engineers, apochrapha has it, always multiplied by π .

Certainly low-Btu gasification and fluidized bed combustion are important and will undoubtedly provide important sources of clean energy in the 1980's. The Environmental Protection Agency will be encouraging substantial R&D effort on these two technologies and will be providing back-up engineering and environmental assessments to ERDA and private efforts. But if we're serious about Project Independence -- and we must be for the sake of our economic security and our independence in foreign relations -- then we must turn to increased use of all of our coals as soon as possible. And to protect public health and welfare, this means using scrubbers on many boilers burning high sulfur coal.

ORD's Stack Gas Scrubbing R&D Program

After arriving in Atlanta last evening, I took a cab downtown in search of a restaurant. Just a few blocks away from the hotel, I noticed a crowd of people entering a church. The sign in front of the church indicated that Mozart's "Requiem in D Minor" was being performed inside. Well, after the recent onslaught of anti-scrubber advertising by American Electric Power, there is little that surprises me but this frankly did. I do find it hard to believe that Donald Cook would have commissioned the performance of Mozart's Requiem in that nearby church just to throw a pallor over this conference. The utter ingenuity of the man!

As I'm sure you're aware, the Environmental Protection Agency views scrubbers much differently from Donald Cook and American Electric Power. And I think you agree or you wouldn't be here at this meeting. Far from being the requiem mass for scrubbers, this gathering is much more a celebration of life. I offer my voice to the choir. Stack gas cleaning of high sulfur coals offers the greatest single contribution, consistent with environmental protection, to the goal of energy self-sufficiency. I feel that if Donald Cook really believed in Project Independence and the need to protect human health from sulfur oxides and metallic acid sulfates he would support the use of scrubbers with high sulfur coal in those parts of the AEP service area where air quality is worse than ambient primary standards.

Problems remain with scrubbers. There has never been a technological innovation which didn't require a shakedown phase. Nuclear power -- the future mainstay of the electric industry -- is still confronting and overcoming problems, 14 to 15 years after

the first civilian demonstration power reactors were started up. We must continue to improve reliability of the non-regenerable (lime/limestone) scrubbers. We must continue to develop improved sludge disposal methods and to evaluate their environmental impacts. We must continue to reduce the capital and operational costs of flue gas desulfurization systems. Finally we must develop regenerable systems which are capable of producing elemental sulfur as a by-product at a reasonable cost.

I'm sure many of you are aware that EPA's Office of Research and Development has come under serious criticism during the past few months. A variety of problems have been pointed out by a National Academy of Sciences panel, the Senate Public Works Committee and others. ORD has already begun to take measures to correct its shortcomings and I'm sure many additional actions will be taken once a new Assistant Administrator is named and confirmed. The major action taken to date is the establishment of a new program management office -- reporting directly to the Assistant Administrator -- which would have responsibility for all of EPA's R&D relating to energy. The Office of Energy Research, of which I'm now serving as Acting Director, is faced with the challenge of meshing EPA's research on health and ecological effects of pollutants from energy plants with EPA's R&D on control technology for those same plants -- an imposing challenge but one which I feel the agency must confront head-on.

One of the features of the partial ORD reorganization is that the Office of Energy Research would concentrate on strategic planning and management, abrogating to the laboratories increased responsibility for detailed project management. This is directly responsive to many of the criticisms. Such a division of responsibility, however, can work only if both parties fully do their jobs. My job begins with setting strategic energy-related goals for EPA's laboratories and I want to begin that job here today with setting goals for the agency's R&D on flue gas desulfurization.

- First, ORD will conduct an advanced lime and limestone test program on our flexible test facility at TVA's Shawnee plant.

This program calls for further-testing and evaluation of the recently demonstrated reliable operating modes and of the "unsaturated gypsum" mode recently discovered in EPA labs. The testing program will also focus on four additional factors: (1) achieving maximum SO₂ removal efficiency consistent with reasonable operating costs, (2) maximizing alkali utilization thus minimizing sludge production, (3) identifying mist eliminator and wash systems with more reliable operation, and (4) evaluating sludge disposal methods, including those offered commercially by Dravo, I.U.C.S., and Chemifix.

- Second, to provide an alternative to lime/limestone throw-away systems, ORD will initiate a full-scale demonstration of a Double Alkali FGD System.

This demonstration will be conducted on a 100-200 MWe coal-fired utility boiler burning high sulfur coal. Selection of the vendor and utility will be done on a competitive basis. ORD will provide up to half of capital and operating costs incurred during the test program.

- Third, to provide an alternative to sulfur-producing synthetic fuel processes, ORD will initiate a full-scale demonstration of a regenerable FGD system which produces by-product sulfur.

This demonstration will also be conducted on 100-200 MWe coal-fired utility boiler burning high sulfur coal. Selection of both the process and the vendor/utility will be done on a competitive basis. Again EPA will cost-share up to half of both capital and operating costs.

Those three goals define the Office of Energy Research's agenda for stack gas scrubbing. We will be doing our best to ensure that both financial and managerial support required to achieve these objectives is provided. We expect ORD's Control Systems Laboratory and other supporting laboratories to move vigorously to meet the challenge. And we invite you to help us meet our obligation to the American public to provide the means for using our domestic coals in environmentally acceptable ways. Now is the time to do it. We can't keep putting off tomorrow. All too quickly tomorrow becomes yesterday.

STATUS OF FLUE GAS
DESULFURIZATION SYSTEMS
IN THE UNITED STATES

Prepared by
Timothy W. Devitt and Fouad K. Zada
PEDCo-Environmental Specialists, Inc.
Suite 13, Atkinson Square
Cincinnati, Ohio 45246

Prepared for Presentation at
Flue Gas Desulfurization Symposium
Sponsored by the Control Systems Laboratory
U.S. Environmental Protection Agency
Atlanta, Georgia

November 4 - 7, 1974

Preceding page blank

ABSTRACT

Data are presented on the number and total MW of flue gas desulfurization (FGD) systems installed, under construction and planned for U.S. utility application. The reliability of operating systems is briefly described. Case histories describe FGD operations at four plants: Hawthorn Power Station (Kansas City Power and Light), Will County Power Station (Commonwealth Edison), Lawrence Power Station (KCPL), and Reid Gardner Power Station (Nevada Power Company). Description of the FGD process at each installation is followed by an account of system performance, with emphasis on problems encountered, solutions developed, and plans for future operations.

CONTENTS

	<u>Page</u>
Status of Flue Gas Desulfurization Technology in the United States	19
The Lawrence Power Station	30
The Will County Power Station	38
Hawthorn Power Station	48
Reid Gardner Power Station	57
Information Sources	62
Appendix A--Flue Gas Desulfurization System Status Report, October 1974	63

STATUS OF FLUE GAS DESULFURIZATION TECHNOLOGY IN THE UNITED STATES

In March 1974, PEDCo-Environmental Specialists was awarded a study by the U.S. Environmental Protection Agency to evaluate the status of flue gas desulfurization (FGD) technology in the United States. This study will have two outputs. The first will be a series of reports on individual FGD facilities. These reports will present values for pertinent process design and operating parameters and describe the operability of the unit from start-up to the time of the plant survey. To date we have visited nine installations; we are scheduled to visit another five facilities by February 1975.

The second study output is a series of monthly reports over the one year study period. These reports include a summary of the operability (reliability) for each utility-size FGD system over the preceding month and a complete list of all FGD units which are either under construction or planned. A copy of the October monthly report is attached as Appendix A.

As shown in Table 1, there are 19 operating FGD systems on utility-size boilers, 17 units under construction and 62 units planned. The planned units include those for which there has been a contract awarded, a letter of intent signed or the utility is presently requesting or evaluating bids. Of the operating systems, 13 are lime or limestone based (including limestone injection), two are magnesium oxide scrubbing, two are sodium carbonate, one is double alkali and one is catalytic oxidation.

Table 1. NUMBER AND TOTAL MW OF FGD SYSTEMS

Status	No. of Units	MW
Operational	19	3,291
Under construction	17	6,777
Planned		
Contract awarded	12	6,640
Letter of intent	4	905
Requesting/evaluating bids	9	3,751
Considering only FGD systems	37	16,472
Total	98	37,836

Figure 1 illustrates the projected increase in the number of megawatts of power production equipped with FGD systems. By 1975, approximately 7000 MW of capacity will be installed of which 50 percent will be retrofit applications and 50 percent will be on new power plants. By 1980, approximately 35,000 MW of capacity will be installed of which 30 and 70 percent will be retrofit and new installations respectively.

Figure 2 illustrates the megawatts of capacity that will be controlled by the FGD process. In 1974, approximately 80 percent is controlled by lime and limestone scrubbing processes; in 1977, over 90 percent will be lime or limestone scrubbing.

Sulfur dioxide removal efficiencies for the operating units range from approximately 70 to 90 percent and particulate removal efficiencies generally are above 99 percent for those units designed for particulate removal. Installation size varies from about 30 MW to over 800 MW. FGD systems have been applied to both low (0.4 to 1.0 percent) and high (6.0 percent) sulfur coals. The non-regenerable systems have used a variety of methods for sludge disposal

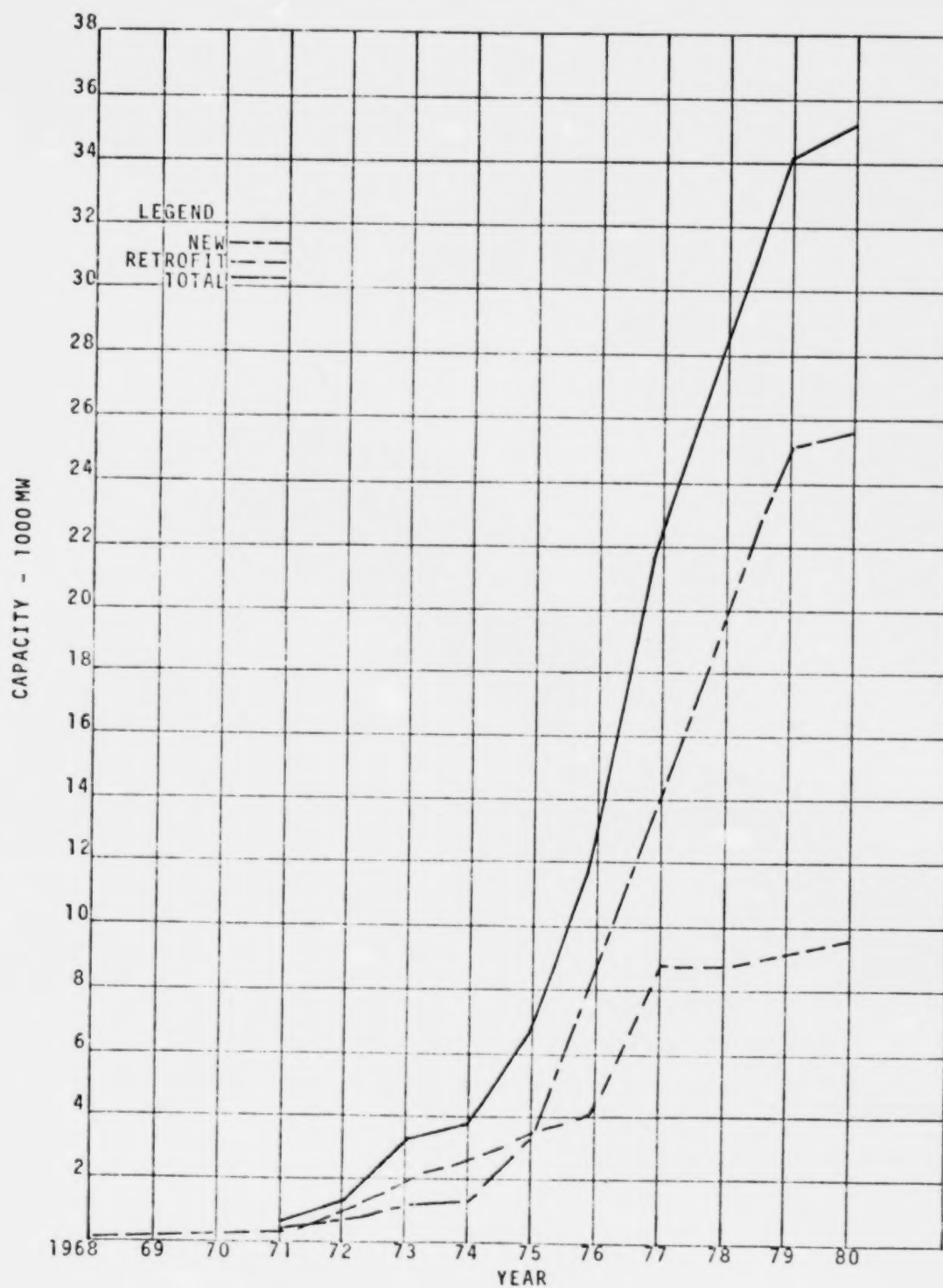


Figure 1. Present and projected FGD systems capacity new/retrofit/total vs. year.

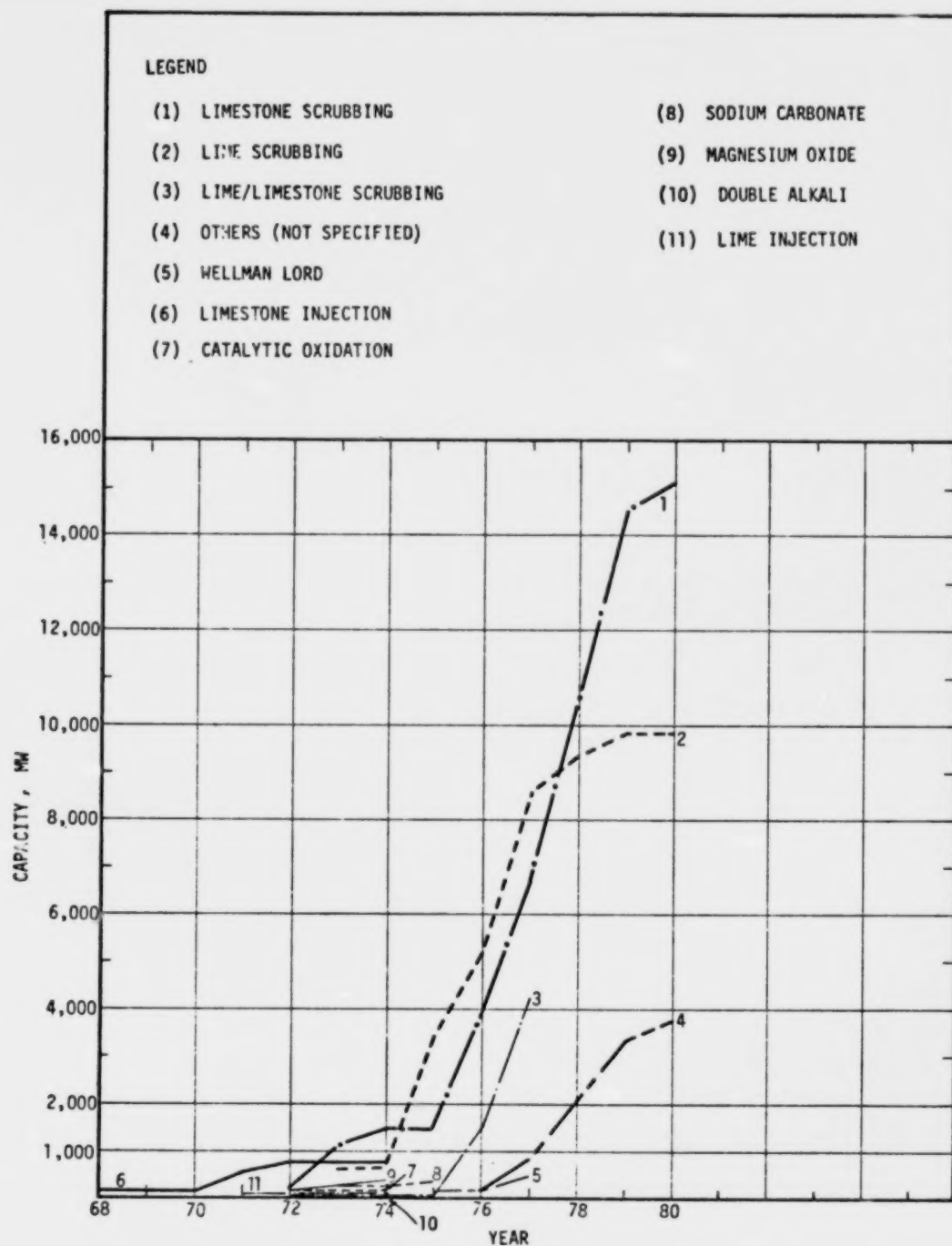


Figure 2. Present and projected capacities of FGD processes.

including ponding, dewatering by vacuum filtration and land-filling, and fixation followed by landfilling. The regenerable systems have been limited to date to recovering sulfuric acid although a sulfur recovery unit is being installed.

The total installed capital costs of these systems have varied from about \$20/KW to over \$100/KW. This cost picture is clouded, however, because of a variety of factors which influenced final installed costs. In some cases, the system designer shared the installation cost and much of the system modification cost with the utility; in other cases there was very little cost sharing. The total installed cost for FGD systems over about 200 MW in capacity is estimated to be in the range of \$60 to \$80/KW.

Pertinent data on the operating systems are presented in Table 2.

The reliability and availability of FGD systems has been a major question. During our plant inspections, it was apparent that most plants had overcome the "chemical" problems; most operators felt they could control the process chemistry and hence minimize or eliminate scaling and related problems. However many units are still faced with a number of mechanical problems (e.g., frozen venturi throat drives, fan vibration). While most of these problems are not major in that they occur on an infrequent basis, they do lead to lower unit availability.

Availability data of any sort and particularly on a common basis are scarce. Some limited data for Paddy's Run and Will County are shown in Figure 3. The availability factor was calculated as hours of FGD system operation divided by hours of boiler operation. Such a definition, however, can be misleading. For example, on a peaking unit like Paddy's Run, the FGD system may be bypassed for short boiler runs even though the system was available for use; thus the availability factor would be lower than the actual

Table 2. DESIGN AND OPERATING DATA FOR OPERATIONAL FGD SYSTEMS

FGD Process/ Power Station	New or Modified	Size of FGD Unit (sq ft)	Status (Date of Start up)	Fuel	Particulate Control			Flue Gas		Sulfur/ SO ₂ Ratio (lb SO ₂ /1000 lb SO ₂)	Water Consumption (gal. per 1000 lb SO ₂)	Reagent Utiliza- tion (%)	FGD System Efficiency		Gas Reheat	Sludge Disposal Method	FGD Costs
				Type	Rate (lb/hr)	Efficiency (%)	Method	Volume (1000 cu ft)	Temp. (°F)				Particulate Removal (%)	SO ₂ Removal (%)			
Altoona, Pennsylvania																	
Altoona Public Service Cholla No. 1	N	115	10/73	Coal	5-15	0.4- 1.0	Yes	No	475,000	263	10	90	99	90	Steam (Indirect)	Pond	\$140/yr
City of Key West Key West Power Plant	N	37	10/72	Oil	1	2.75	Yes	No	770,000	355	15	71	95	88	Steam (Indirect)	Pond	\$22.4/yr
Comstock Edison Will County No. 1	N	167	2/73	Coal	10.0	2.0	No	Yes	770,000	355	15	90	95	90	Steam (Indirect)	Pond	\$43/yr
Kansas City Power & Light La Crosse No. 1	N	420	6/73	Coal	20-28	6.0	No	No	2,300,000	328	18	90	95	90	Steam (Indirect)	Pond	\$70-75/yr
Altoona, Pennsylvania																	
Altoona Public Service Phillips	N	410	10/73	Coal	18	2.3	Yes	Yes	540,000	342	NA	90	99.3	80	Steam (Indirect)	Pond	\$15-18/yr
Louisville Gas & Electric Perry No. 6	N	45	6/73	Coal	14	3.7	No	Yes	500,000	330	20	100	99	80	Gas Fired Spitzer Direct	Vacuum Fil- ter & Reel	\$57/yr
Southern California Edwards Nobles No. 2	N	140	11/73	Coal	0.5- 0.8	0.8	No	Yes	2,300,000	NA	83	NA	-	-	Not Reheat	Pond	-
Altoona, Pennsylvania																	
Kansas City Power & Light Northern No. 4	N	100	6/72	Coal	14	3.75	No	No	445,000	280	21-26	NA	99	70	Steam (Indirect)	Pond	\$19.0/yr
Kansas City Power & Light Northern No. 3	N	140	11/72	Coal	14	3.75	No	No	445,000	300	21-26	NA	99	70	Steam (Indirect)	Pond	\$19.6/yr
Kansas Power & Light Lawrence No. 5	N	480	11/72	Coal	12.5	1.4	No	No	990,000	300	23	62	98	75	Not Reheat	Pond	NA
Kansas Power & Light Lawrence No. 4	N	135	12/48	Coal	12.5	1.4	No	No	338,800	300	22	62	99	75	Not Reheat	Pond	NA
Altoona, Pennsylvania																	
Delaware Power - Comp Alma Station	N	80	6/71	Coal	12-18	3.0- 3.5	Yes	No	-	310	-	NA	-	70	Not Applicable	Pond	NA
Altoona, Pennsylvania																	
Thames Valley Thames No. 10	N	30	4/73	Coal	12-15	3.0	No	No	10,000	300	64	66	95+	80-95	Steam (Indirect)	Pond	NA
Altoona, Pennsylvania																	
Waco Electric & Power Bickerman No. 3	N	150	6/72	Oil	-	2.8	No	Yes	440,000	300	NA	100	-	90	Not Reheat	Acid Plant	\$25/yr
Altoona, Pennsylvania																	
Sodium Chloride Sodium Chloride	N	108	9/72	Coal	20	3.0	Yes	Yes	395,000	239	NA	100	99.5	90	Acid Plant	Acid Plant	\$70/yr
Altoona, Pennsylvania																	
Merced Power Merced No. 2	N	135	12/73	Coal	9	0.4	Yes	No	473,000	350	10	100	99	95	Direct Not Reheat	Pond or Dry Lake	NA
Merced Power Merced No. 1	N	135	12/73	Coal	9	0.4	Yes	No	473,000	350	10	100	99	95	Direct Not Reheat	Pond or Dry Lake	NA
Altoona, Pennsylvania																	
General Motors General Motors No. 1, 2, 3 and 4	N	30	3/74	Coal	NA	NA	Yes	No	NA	NA	NA	NA	NA	NA	NA	Pond	\$75/yr
Altoona, Pennsylvania																	
Illinois Power Illinois No. 4	N	110	10/73	Coal	10	3.5	Yes	Yes	1,120,000	310	-	-	97	88	NA	Acid Plant	\$71/yr

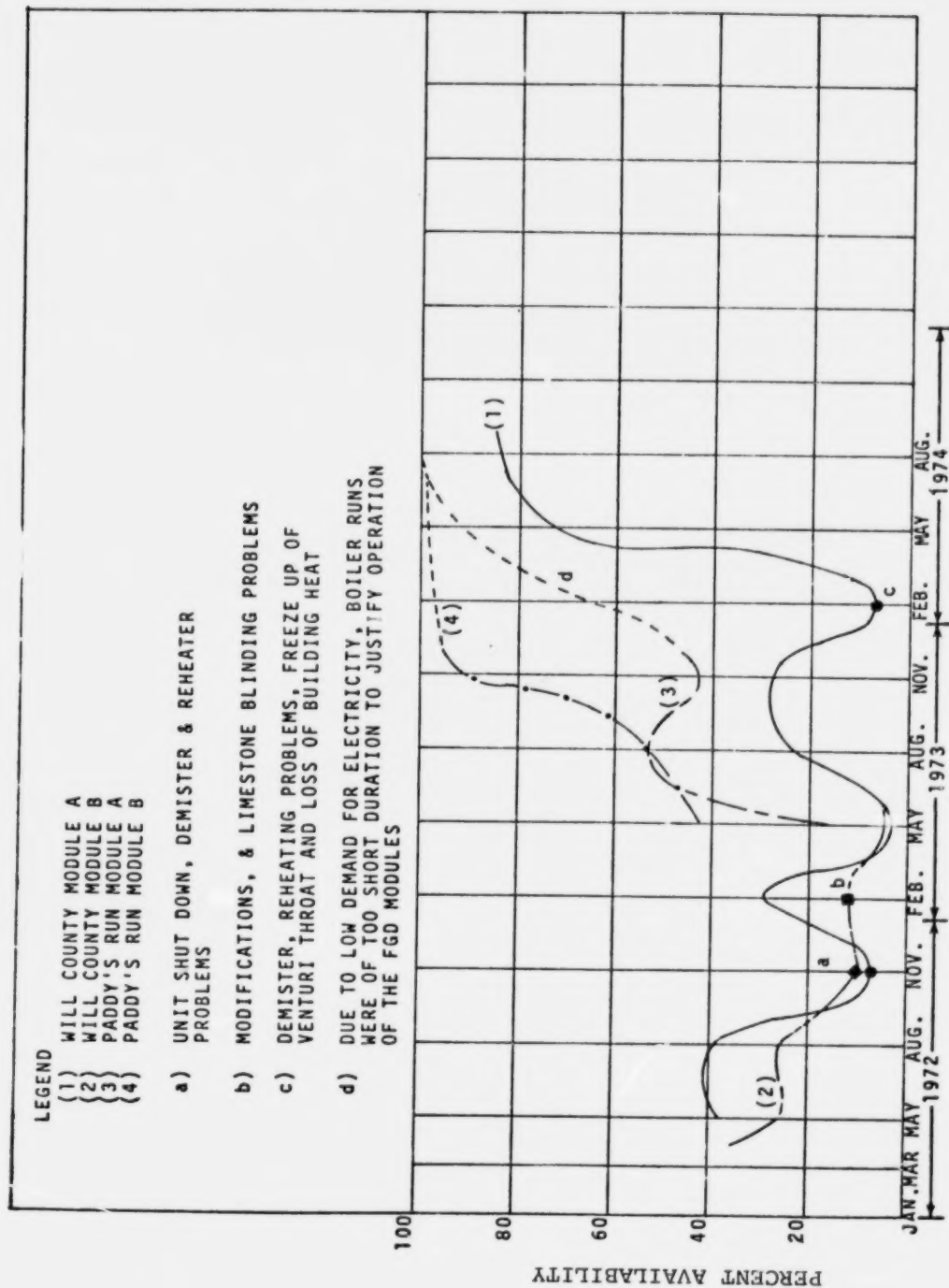


Figure 3. FGD system availability.

availability of the unit. On the other hand, units with no bypass capability would have a high availability factor since the boiler would be down when the FGD system was inoperative. Since complete data were not available and could be misleading if not adequately qualified, we have presented a qualitative description of recent performance for all the operating systems in Table 3.

The FGD systems at the Lawrence, Will County, Hawthorn and Reid Gardner installations are described in the following sections. These systems were selected since they are not covered in other papers presented at this symposium.

Table 3. OPERATING SUMMARY FOR FGD SYSTEMS

Utility and Unit Name	Status
Arizona Public Service Cholla No. 1	Since start-up in October 1973, the FGD system availability has been consistently above 90 percent. There have been a few mechanical problems, the most persistent being vibration in the reheater section.
Boston Edison Mystic No. 6	Scrubber availability reported for 1973 ranged from 73 percent in August to 13 percent in December. The decreasing availability was due to deterioration and subsequent overhaul of process equipment. Recent availability figures are: March 87 percent, April 81 percent, May 57 percent and June 40 percent. This system is down with no immediate plans for restarting since the EPA demonstration program is completed. Boston Edison is evaluating the data collected during the demonstration period to determine the course of future action.
Commonwealth Edison Will County No. 1	Availability of Module A has increased in the past several months. It was 72 percent in April, 93 percent in May, 54 percent in June, 95 percent in July, 91 percent in August and 85 percent in September. Module B was removed from service until all necessary modifications are made to Module A to permit reliable operation.
Dairy Power Co-Op Aima Station	This is a demonstration project where lime mixed with water is injected into the boiler. The longest test run was less than two days.
Duquesne Light Phillips Station	At present, only flue gas from Boilers 2, 3 and 4 (about 40 percent of the station capacity) is treated because fly ash from the inefficient precipitators overloads the clarifiers. Operating hours for Modules 1 through 4 between 3/17/74 and 6/30/74 were 1756, 762, 815, 1707 respectively. The plant tries to run Modules 1 and 4 continuously. The outage hours for these units is primarily for inspections.
General Motors Parma Plant	System availability has been essentially 100 percent since 4/74. However the availability figure has little meaning since only two of the four modules were operating at any time because of low demand.
Illinois Power Wood River	The unit operated only 700 hours during the past two years. Primary reason for the downtime was the delay in converting the reheaters from natural gas to fuel oil firing. The unit is scheduled to restart again in late November.

Table 3 (continued). OPERATING SUMMARY FOR FGD SYSTEMS

Utility and Unit Name	Status
Kansas City Power and Light Hawthorn No. 3	The FGD system has undergone several major modifications. Availability has since increased from 30 percent in 1973 to as high as 70 percent recently. Plant operators were on strike between July and October 1, 1974, and during that time the FGD system was down. The system was restarted on October 1st.
Kansas City Power and Light Hawthorn No. 4	This unit was converted from furnace injection to tail-end scrubbing. The plant has experienced more problems with this unit (both mechanical and chemical in nature) than with Unit 3. The operation can be characterized as "fair."
Kansas City Power and Light La Cygne	Initial problems included fan deposits, demister and nozzle pluggage, reheater failure and screen pluggage. Many of the original problems were attributed to poor pH control. Recently the unit has had about 80 percent availability and there are no outstanding problems. However it is necessary to remove each module from service once a week to clean out accumulated solids.
Kansas Power and Light Lawrence No. 4	The system has undergone a number of process modifications. At present the system is capable of sustained operation although SO ₂ removal efficiency is limited to only 75 percent and the demisters must be washed daily (automatic washing). Manual washing of the demisters is required every two weeks. The unit is badly corroded and will be replaced by a new electrostatic precipitator and FGD system in 1977.
Kansas Power and Light Lawrence No. 5	This unit has encountered many of the same process problems as Unit 4. In addition there is poor gas flow distribution between the eight modules and within each module; measures are being taken to correct the distribution. The boiler burns natural gas whenever it is available so it is difficult to assess FGD system availability.
Louisville Gas and Electric Paddy's Run	FGD system availability is near 100 percent. However since the system is installed on a peaking boiler there are many occasions when the boiler's run is too short to justify start-up of the FGD system.
Southern California Edison Mohave No. 2	This is an experimental unit. It has operated quite successfully with an availability above 80 percent.

Table 3 (continued). OPERATING SUMMARY FOR FGD SYSTEMS

Utility and Unit Name	Status
Nevada Power Reid Gardner No. 1	The FGD system started-up in March 1974 and has had an availability of over 90 percent when there was sufficient trona (impure form of sodium carbonate). The lack of trona has limited the operation of this unit. The system has operated only 900 hours since April 1974.
Nevada Power Reid Gardner No. 2	The FGD system started-up in March 1974 and has had an availability of over 90 percent when there was sufficient trona (impure form of sodium carbonate). The lack of trona has limited the operation of the unit. The system has operated only 900 hours since April 1974.
Potomac Electric and Power Dickerson No. 3	There is no record of FGD system's availability for the period prior to August 1974 since the unit was down frequently for process modification, equipment repairs, etc. Availability since August is about 34 percent.

THE LAWRENCE POWER STATION

Power Generating Facilities

The Lawrence Power Station of Kansas Power and Light Company is located in a lightly industrialized area on the outskirts of Lawrence, Kansas.

The plant operates two steam boilers which are equipped to burn coal, natural gas supplemented with oil, or a combination of these three fuels. Boiler 4 is the oldest of the two units. It was first placed in service in 1959 and operated as a cyclic-load boiler. The maximum electric-generating capacity of this unit varies with the type of fuel being burned: with natural gas the output can be as high as 143 MW; with coal plus natural gas output decreases to 125 MW. Retrofitting of this boiler with an FGD system in 1968 introduced additional pressure drop in the flue gas system and further reduced the boiler capacity to 115 MW.

The second unit at the plant is Boiler 5, with a rated capacity of 400 MW when burning coal plus natural gas. The unit together with the FGD system, was placed in service in 1971. Like Boiler 4, it is classified as a cyclic-load unit.

Both boilers at the Lawrence Power Station were built by Combustion Engineering, which also designed and installed the FGD systems on these boilers. These FGD systems consist of limestone furnace injection with flue gas wet scrubbing.

The present grades of coal burned at the Lawrence Power Station has a gross heat content of 12,000 BTU/lb. Average ash

and sulfur contents are 12 percent and 3.75 percent, respectively. The company is in the process of switching from this high-sulfur Kansas coal to Wyoming coal which contains 0.4 to 0.8 percent sulfur and 10 percent ash, with a gross heating value of 10,000 BTU/lb. This change was necessitated by the curtailment of strip mining operations at the Kansas coal supply site.

FLUE GAS DESULFURIZATION SYSTEM

FGD Process Description. The flue gas desulfurization systems on Boilers 4 and 5 are identical in basic design and operation. The FGD system on Boiler 4 has undergone several major modifications since its start-up in November 1968, many of which were later incorporated into the design of the FGD system on Boiler 5.

The FGD system associated with each boiler incorporates facilities for storing, pulverizing, and injection of finely ground limestone rock into the boiler's furnace chamber, where the bulk of it is calcined. The calcined limestone, along with the fly ash, is transported by the flue gas to the tail-end wet scrubber modules, where the SO_2 in the gas reacts with the scrubbed lime/limestone in a recirculated slurry and is substantially removed, along with the fly ash, from the gas stream. The cleaned gas is then demisted and reheated (to prevent condensation in the downstream equipment) and finally discharged from the stack by induced-draft fans.

There are two FGD modules on Boiler 4 and eight FGD modules on Boiler 5. The 10 modules are identical in size, and each is designed to handle roughly 150,000 scfm of flue gas. A typical module is shown in Figure 4. A single stage of 3/4-inch-diameter glass marbles is contained in a bed about 3 to 4 inches thick, fitted with overflow pots to collect and drain the liquor from the top of the bed. The scrubbing liquor is sprayed through nozzles located below the bed.

Each of the two modules on Boiler 4 is connected (through an I.D. fan) to a separate 120-foot stack; gases from all eight

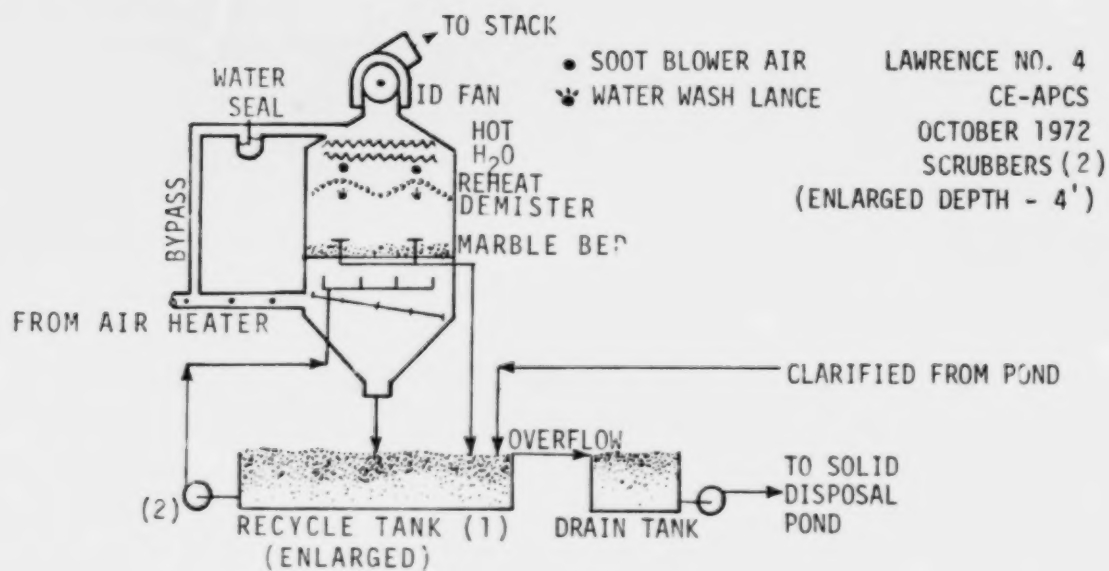


Figure 4. Sketch of a typical FGD module at the Lawrence Power Station.

Source: Kansas Power and Light Company.

modules on Boiler 5 are discharged through a common stack, 375 feet tall.

Originally, all modules were fitted with bypass ducts and hydraulic seal dampers. Because of extensive corrosion and pluggage problems with the two modules of Boiler 4, the bypass ducts on these modules were removed.

Spent liquor from the scrubber tower drains into a recycle tank, which allows 30 to 40 minutes retention time to ensure complete conversion of the scrubbed SO_2 to calcium sulfite and calcium sulfate. Spent liquor from this tank overflows to a drain tank, from which it is pumped to sludge disposal ponds.

There are three sludge ponds on site, covering areas of 4 acres, 16 acres, and 28 acres. The sludge first enters the 16-acre pond and overflows into the 4- and 28-acre ponds. Approximately 800 gpm of sludge containing 9 percent solids, is fed to the unlined ponds. Because the sludge contains unreacted lime as well as fly ash (ingredients which are usually added to stabilize limestone sludge), it sets up very hard, like concrete, without additives.

FGD System Performance. Several major modifications completed by Combustion Engineering and Kansas Power and Light Company on Unit 4 appear to have significantly improved the performance and availability of that unit. Both FGD modules on Boiler 4 have recently performed reasonably well. Availabilities close to 100 percent were reported for July and August 1974. The SO_2 removal efficiency has been around 65 percent, which is sufficient for the plant to comply with the applicable pollution control regulation. SO_2 removal efficiencies as high as 85 percent were achieved over a short period, but only at the expense of an accelerated rate of scale formation in the scrubbers.

Because recent operation of Boiler 5 has been on natural gas, the FGD modules have not been subjected lately to severe operating conditions. When the boiler is coal-fired and the

FGD system is in operation, problems experienced are similar to those encountered with the modules on Boiler 4. Currently the main problem with Unit 5 is improper distribution of flue gas to the eight modules. Combustion Engineering is performing tests on Unit 5 to alleviate this problem.

Analysis of the problems encountered during and since start-up reveals that nearly all were due to improper control of process chemistry. In limestone furnace injection, it is difficult to achieve satisfactory control of the degree of limestone calcination as well as the amount of lime/limestone carried in the flue gas to the tail-end scrubbers. This situation is further aggravated when the boiler is operating as a cyclic-load boiler and is fired with variable combinations of coal, natural gas, and oil.

As mentioned earlier, the FGD system on Boiler 4 underwent several major modifications which are briefly described in this section.

When the FGD system initially started operation in 1968, the configuration of each of the two modules was as shown in Figure 5. This design presented many operating problems, which included (1) scale buildup and plugging of the hot gas inlet duct, (2) erosion of the scrubber walls and corrosion of the scrubber internals, (3) plugging and scaling of drain lines, tanks, pumps, marble bed, demister, and reheater and (4) scale build-up on I.D. fan rotors, which resulted in fan imbalance and vibration.

In addition to the problems just mentioned, the SO_2 removal was quite low because of overburning of limestone in the furnace and dropout of the lime with the ash in the bottom of the scrubbers.

After the first few months of operation, the scrubbers were modified, as shown in Figure 6, to include (1) addition of soot blowers in the gas inlet duct and under the reheater

LAWRENCE NO. 4
CE-APCS
DECEMBER 1968
SCRUBBERS(2)

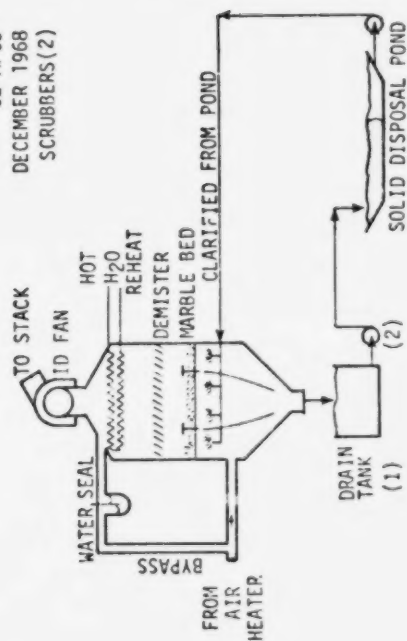


Figure 5.

LAWRENCE NO. 4
CE-APCS
OCTOBER 1970
SCRUBBERS(2)
• SOOT BLOWER AIR

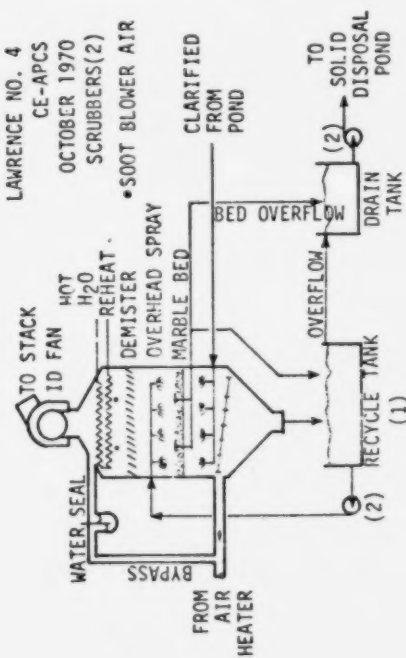


Figure 7.

LAWRENCE NO. 4
CE-APCS
OCTOBER 1969
SCRUBBERS(2)
• SOOT BLOWER AIR

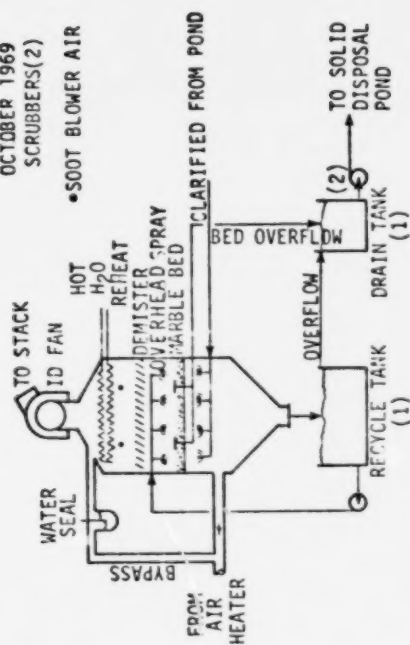


Figure 6.

LAWRENCE NO. 4
CE-APCS
OCTOBER 1972
SCRUBBERS(2)
(ENLARGED DEPTH 4')
• SOOT BLOWER AIR
• WATER WASH LANCE

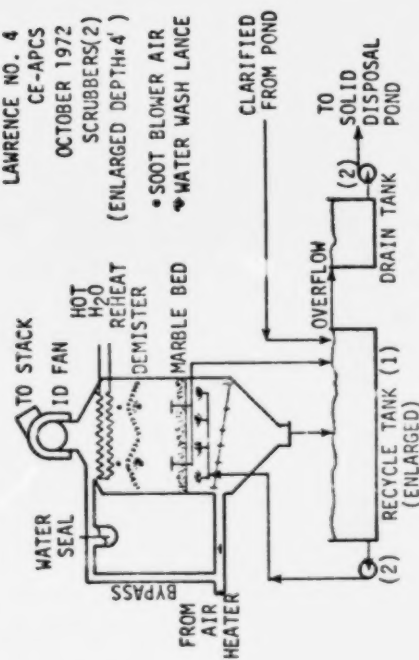


Figure 8.

bundle to prevent plugging (2) raising of the demister to reduce carryover of solids, (3) directing the overflow liquor from the pots to the pond, and (4) installation of a large recycle tank and pump to catch the highly alkaline underflow and recirculate it to the marble bed. Other modifications to combat corrosion and pluggage were installation of a new type of spray nozzles and lining of the bottom section of the scrubber tanks with gunite.

Most of the problems were reduced but not eliminated by these modifications. The new recirculation system did improve the SO₂ removal efficiency of the scrubbers.

In an effort to further minimize corrosion, erosion, scaling, and plugging, additional revisions were made during the summer of 1970. The configuration of the scrubber in October 1970 is shown in Figure 7. The major revisions were the following:

1. Sandblasting and coating the interior of the scrubbers with two coats of glass flake lining.
2. Replacement of all internal steel pipes with plastic and fiberglass.
3. Replacement of the stainless steel demisters with fiberglass demisters.
4. Addition of a ladder vane under the marble beds to improve gas distribution.
5. Modification of the pot overflow drain piping to allow the liquor to return to the recycle tank for a semiclosed liquor loop operation.
6. Removal of the original copper fin tubes of the reheater coils, which because of close fin spacing, caused the reheaters to plug easily. Also, the fins were flattened by the soot blower jets. The copper units were replaced with a carbon steel tube fin coil.

Demister pluggage continued to create serious problems. Manual washing every other night was required to maintain the output required of the unit.

In the summer of 1972 the scrubbers (on both Units 4 and 5) were modified to operate using a high solid slurry crystallization process to control saturation and precipitation of scale within the scrubber. These latest major modifications, shown in Figure 8, included enlargement of the liquor recirculation tank and replacement of many components, such as piping, nozzles, pumps, and mixers. Also, the demisters were replaced with a new two-bank fiberglass unit fitted with high-pressure wash-water lances.

Operation of the two FGD systems since the fall of 1973 has been most successful to date. Some of the remaining problems are isolated areas of corrosion, failure of expansion joints, fouling of the demister, rapid wear of slurry pumps, and failure of valves.

Future modifications of the FGD systems on each boiler will be concerned primarily with alleviating problems inherent in furnace injection with limestone. Therefore, Unit 5 will be converted to a tail-end, wet limestone scrubbing process by fall 1975.

The current schedule for revisions on Unit 4 is as follows:

- a) By January 1975, construction of two 2-stage scrubbers (venturi followed by spray) will be started.
- b) By January 1976 the two new scrubber modules will be operational. The present scrubbers will be kept in service while the new scrubbers are being built.
- c) By January 1977 the present scrubbers will be razed and an ESP will be installed. It is anticipated that the ESP/venturi/spray flue gas cleaning system will be operational by January 1977. The new system will have forced oxidation via aeration to produce sulfates.

THE WILL COUNTY POWER STATION

Power Generating Facilities

The Will County Power Station of Commonwealth Edison Company is located on the Chicago Sanitary and Ship Canal near the town of Romeoville, in Will County, Illinois. Delivery of coal and limestone to the Will County Power Station is primarily by barges using this canal.

The station has four electric-power generating units with a total rated capacity of 1147 MW. Only Unit 1 is retrofitted with a flue gas desulfurization system.

Unit 1 has a wet-bottom coal-fired boiler producing 167 MW of electricity. The boiler was manufactured by Babcock and Wilcox and was installed in 1935.

The coal presently being burned has an average gross heating value of 9463 BTU/lb; ash and sulfur contents are 10 and 2.1 percent, respectively.

The boiler is fitted with an electrostatic precipitator, having a 79 percent actual particulate collection efficiency. Normally, the precipitator is not used except when the FGD scrubbers are out of service.

FLUE GAS DESULFURIZATION SYSTEM

FGD Process Description. The wet limestone flue gas desulfurization system was placed in service on February 23, 1972. The system consists of limestone handling and milling facilities, two FGD scrubber modules (identified as A and B), and a sludge treatment and stabilization unit. Each FGD module consists of a booster fan and a Venturi scrubber followed by a turbulent

contact scrubber (TCA) SO_2 absorber tower. Operational problems, described later in detail, have caused shutdown of Module B and modification of Module A.

The FGD process is described in terms of the three major operations: limestone milling, particulate and SO_2 scrubbing, and sludge treatment and disposal.

The limestone milling system consists of a limestone rock conveyor, two 260-ton-capacity limestone storage silos, two wet ball mills, and a slurry storage tank. The total storage capacity of the two silos is equivalent to 48 hours of the FGD requirement of limestone at full load. The limestone is 97.5 percent calcium carbonate, 0.99 percent magnesium carbonate, and 0.48 percent silica. It is received as coarsely ground particles (about 1/2 inch diameter or less) and is finely ground to 95 percent through 320 mesh. Grinding is accomplished in two wet ball mills, each rated at 12 tons per hour. The limestone is discharged from the mills in the form of a water slurry containing 20 percent solids. The slurry is piped to a 4-hour capacity, 62,500-gallon storage tank, which supplies the limestone for the scrubber modules.

The two scrubber modules on Will County Boiler 1 are shown in Figure 9. Each module was designed for 385,000 acfm throughput at 355°F, each handling 50 percent of the 167-MW boiler's flue gas load. The liquid and gas flow through a typical module is shown in Figure 10. The flue gas emerges from the boiler, passes through an electrostatic precipitator and is directed to the venturi scrubber. The gas is forced through the rectangular venturi throat and comes in contact with jets of slurry sprayed from high-pressure nozzles located on each side of the throat. The scrubbing efficiency and consequently the pressure drop in the venturi are regulated by actuating a drive mechanism that varies the width of the venturi gap.

The quenched flue gas mixed with slurry droplets exits from the venturi throat and turns through the sump, in which

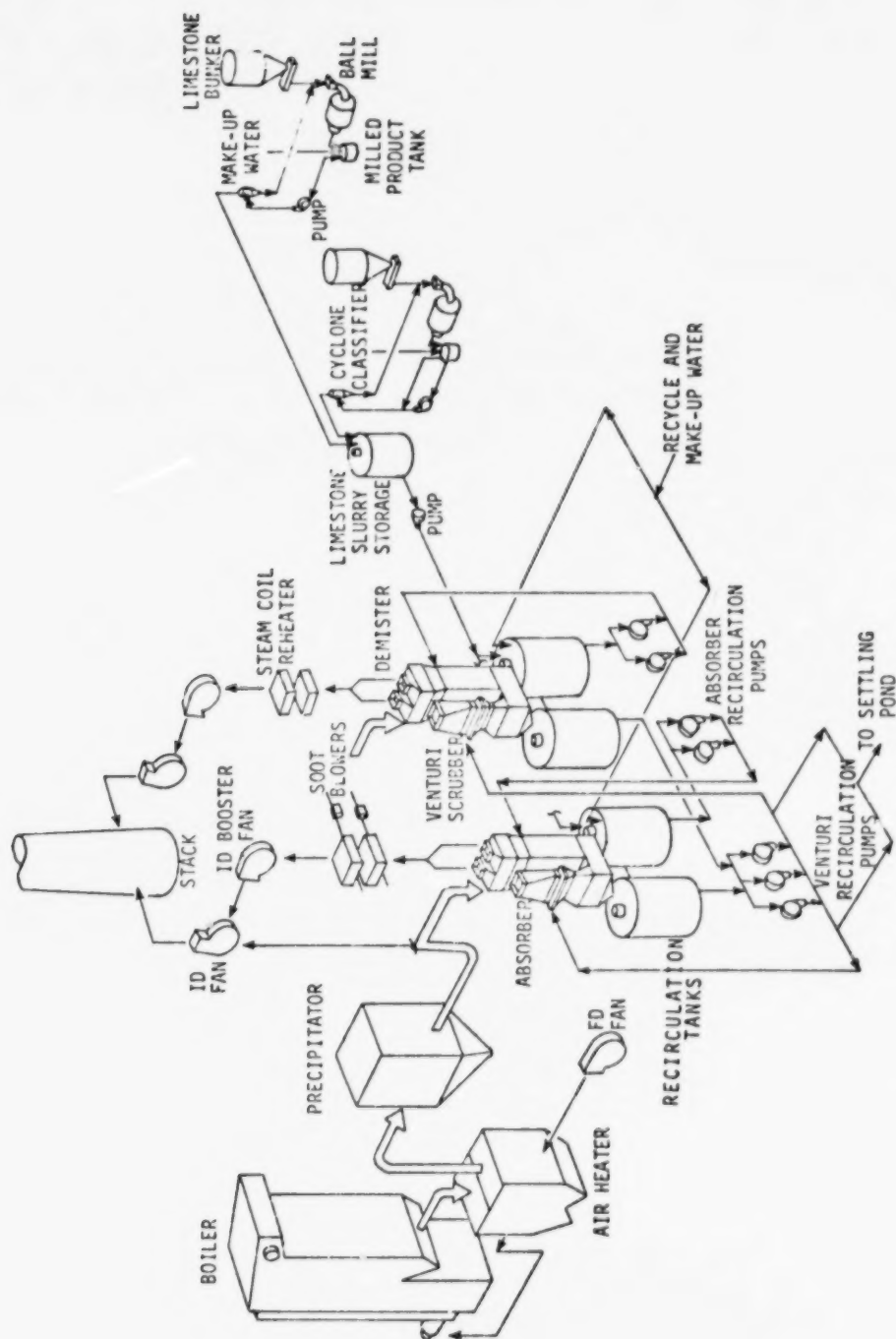


Figure 9. General flow diagram of the FGD system at Will County Unit 1.

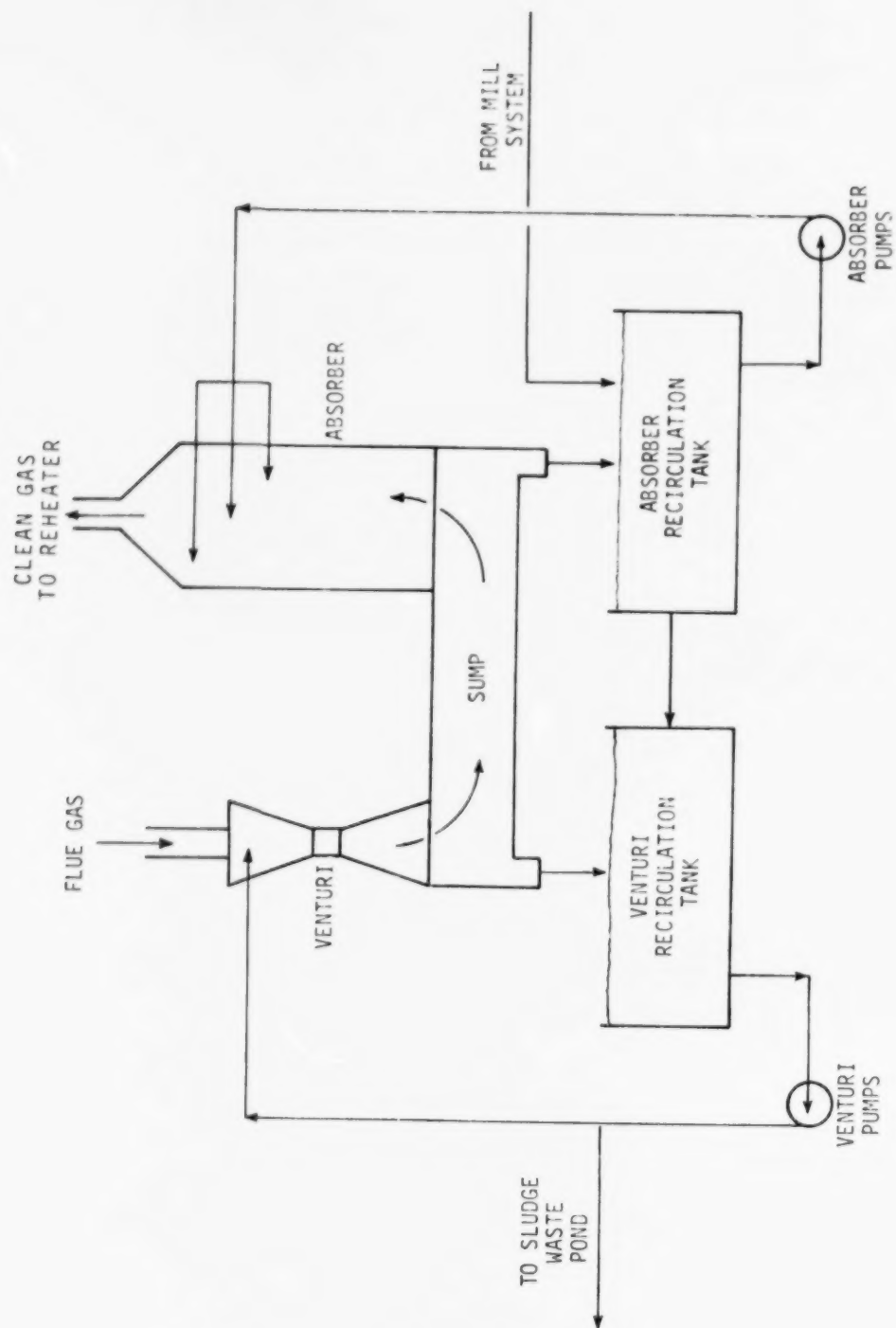


Figure 10. Flow diagram of a typical FGD module.

a great reduction in flow velocity causes the slurry droplets to fall out of the gas stream into the venturi circulation tank.

The gas then flows upward through the SO_2 scrubber tower, and passes through two perforated trays, which are wetted with a shower of limestone slurry circulated from the SO_2 scrubber tank. These trays provide an extensive contact surface for absorption of SO_2 from the circulated slurry.

The scrubbed gas, now essentially free of fly ash and SO_2 , rises up the tower and passes through a two-stage Chevron-type demister. The fine mist droplets of liquid carried over with the gas coalesce on the demister's vanes, and the resulting large droplets fall through the tower by gravity. The demister is equipped with two sets of wash water lances. It is washed continuously from below by 120 gpm of fresh water and also intermittently from above by 1000 gpm of pond water for 30 seconds every 2 hours. The demisted gas then enters the reheater unit, where its temperature is raised from 128°F to about 200°F. This temperature increase imparts buoyancy to the gas and prevents condensation in the fans, ducts, and the brick-lined steel stack.

The bare-tube reheater is divided into three sections; the first is made of 304 stainless steel and the other two of corten steel. Each reheater has four soot blowers to maintain tube cleanliness. The heat is supplied from Unit 1 boiler as 350 psig, saturated steam.

To compensate for draft loss across each scrubber module, an induced-draft fan boosts the pressure of the heated gas and delivers it to the suction side of the boiler induced-draft fan.

There are two slurry tanks for each module: the SO_2 scrubber circulation tanks, to which fresh limestone and make-up water are added, and the venturi scrubber circulation tank, from which the spent limestone slurry (or sludge) is discharged. These two tanks are interconnected in such a way that the spent liquor

from the SO₂ scrubber tank overflows into the venturi circulation tank. Each tank is fitted with agitators and pumps. The slurry recirculation rate in the SO₂ scrubber is about 8750 gpm. The recirculation rate through the venturi is 5800 gpm.

The sludge treatment and disposal system used at Will County is illustrated in Figure 11. Spent slurry from all the venturi tanks is discharged to a 65-foot-diameter clarifier tank. During emergencies or when the clarifier or stabilizer unit is inoperable, the slurry can be discharged to a small clay-lined holding pond. Overflow from the clarifier is returned to the process; underflow is stabilized in a facility built and operated by Chicago Flyash Company under contract with Commonwealth. About 200 pounds of lime and 400 pounds of fly ash are used per ton of dry solids of sludge. The fixed sludge is transported by concrete mixing trucks to a small pond on site for solidification. The stabilized sludge solidifies in about 1 week, after which it is hauled by Material Services Company to an offsite dump for use as landfill. Recently, this contractor obtained State-EPA permit to develop an offsite land for sludge disposal.

FGD System Performance. As mentioned earlier, the FGD modules at Will County 1 were plagued with numerous problems since February 1972. Many of these problems have since been solved, and others have become manageable. Improved operation of the FGD system is reflected in increasingly higher availability figures for Module A in 1974. When Module B is modified on the basis of experience gained with Module A, its performance and availability should be comparable to those obtained from Module A to date. The monthly availability factors for each module are presented in Table 4. A brief chronology of the major problems encountered and their solutions is given below.

1972 - Demister plugging was a constant problem, mainly because of heavy accumulations of limestone on the bottom of the demister, which were partly due to low wash-water pressure

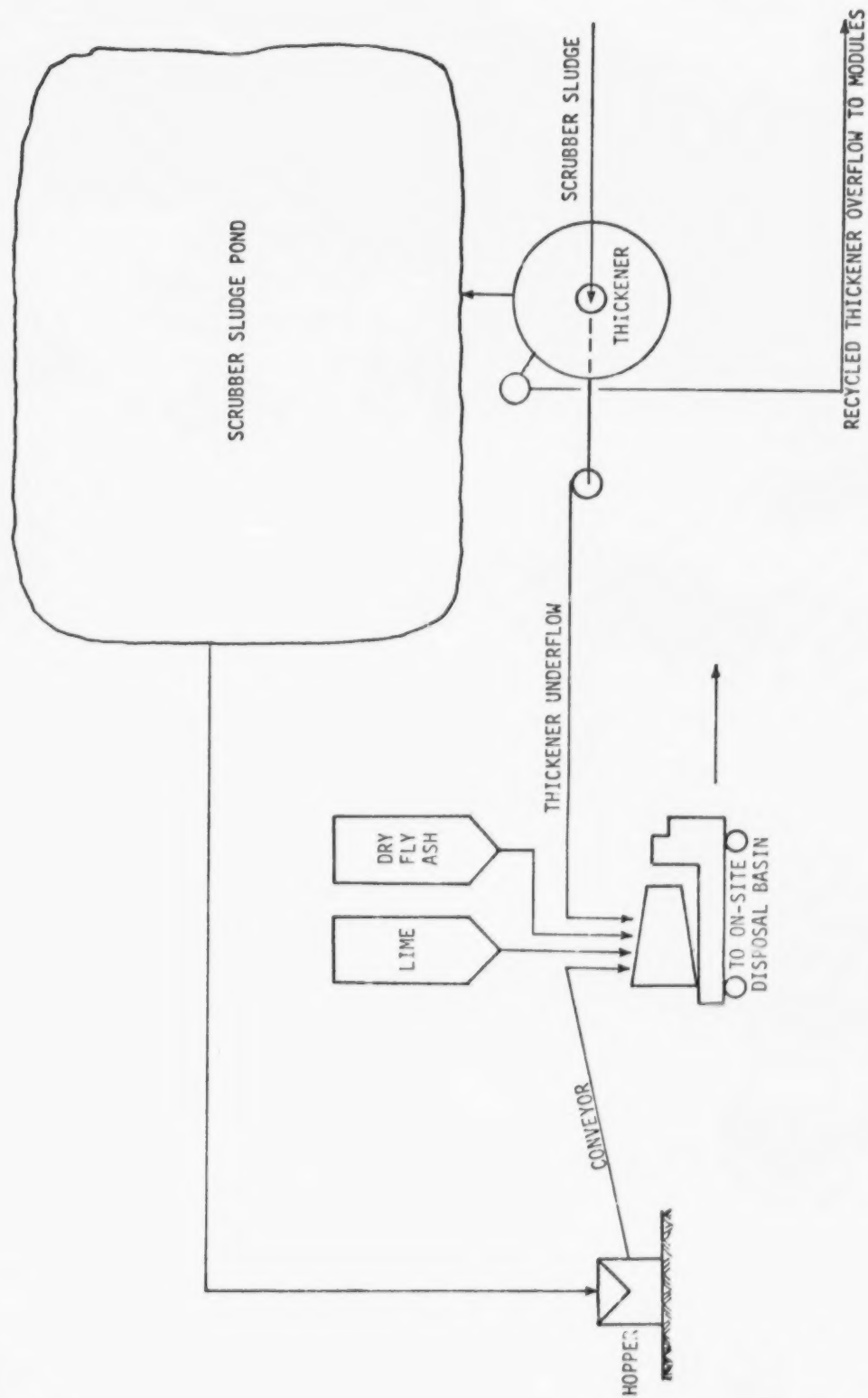


Figure 11. Limestone sludge stabilization facilities.

Table 4. FGD SYSTEM AVAILABILITY FACTORS

FOR WILL COUNTY UNIT NO. 1

Period Month/year	Availability, %		Period		Availability, %		Period		Availability, %	
	Module A	Module B	Month/year	Module A	Module B	Month/year	Module A	Module B	Month/year	Module A
March, 1972	0	35.4	January, 1973	0	0	January, 1974	0	0	0	0
April	33.99	13.68	February	21.94	24.26	February	0	0	0	0
May	69.48	31.82	March	64.79	10.69	March	20.98	0	0	0
June	8.44	30.91	April	6.17	13.08	April	72.27	0	0	0
July			May	0	0	May	93.12	0	0	0
August	78.65	20.60	June	0.96	0	June	54.50	0	0	0
September	0	29.48	July	51.36	0	July	95.75	0	0	0
October	0	0	August	19.23	0	August	91.25	0	0	0
November	0	0	September	0	0					
December	21.77	29.69	October	32.00	0					
			November	50.80	0					
			December	0	0					

caused by leaks in the pond-return bypass. Because of plugging, Modules A and B were out of service for several days each month during March, April, June, and July 1972. The modules were also out of service from September 26 to November 21, 1972, because the boiler was down during that period.

As a means of solving the demister problem, the scrubbers' slurry nozzles were lowered and the slurry circulation system was left out of service to keep the demister clean. Because this change gave no improvement, the demister elements were hand-washed, a procedure that decreased demister plugging but increased difficulties with the venturi nozzle because broken elements from the demister entered the slurry system.

Because the reheater of Module B vibrated excessively during high rates of gas flow, Module B was taken out of service in April for reheater modifications. These modifications included rebracing of the reheater tubes and installation of a baffle plate to reduce vibrations.

Other reasons for outages in 1972 were erosion and plugging of spray nozzles, internal and external buildup of deposits on venturi nozzles, corrosion cracking, sulfite blinding, and fan vibrations.

1973 - Demister plugging continued in 1973. Furthermore, the demister on Module B broke loose from its mountings and the resultant carryover of wash water plugged the reheater, also leaked because of chloride pitting corrosion. Module A was down from April 24 to May 24, 1973, and Module B has been down from April to date. No scrubbers were in operation from August 27 to September 26, 1973.

As a further step toward resolving the pluggage problems, the system of continuous underspray and intermittent overspray was installed in Module A to wash all the demister compartments. Extra nozzles were added and a clean water supply was maintained. The reheater bundle was also retubed.

1974 - This year to date only Module A is in operation. Although plugging of the demister and reheater has not been completely eliminated, the improved washing system has provided a partial solution. The other significant problems are freeze-up of the venturi throat drive mechanism, tank screen blinding, duct corrosion, and vibrations. Loss of building heat caused extensive damage and kept the scrubber down the entire month of January 1974.

Some components of Module B, such as reheater bundles have been cannibalized for use in Module A. Modifications of Module B are not expected to begin until performance of Module A is satisfactory.

HAWTHORN POWER STATION

Power Generating Facilities

The Hawthorn Power Station of Kansas City Power and Light Company is located in a heavily industrialized area on the north bank of the Missouri River in East Kansas City, Missouri.

The plant operates five coal-fired boilers. Boilers 1 and 2 are considered peak boilers, each rated at 100 MW. Boilers 3 and 4 are cyclic boilers. When burning natural gas, each is rated at 140 MW; when burning coal the capacity of each boiler drops to 100 MW. These four boilers were built and placed in service in early 1950. Boiler 5 was placed in service in early 1970; it operates as a base-load boiler at a rated capacity of 500 MW. All these boilers are dry-bottom, pulverized-coal-fired units, designed and manufactured by Combustion Engineering.

Two grades of coal are burned: the coal with higher ash content typically contains 14 percent ash and 3 percent sulfur, with a heat content of 11,400 BTU/lb; the lower-ash coal contains 11 percent ash and 0.6 percent sulfur, with a heat content of 9800 BTU/lb. Natural gas is also burned occasionally.

Of the five boilers at the Hawthorn plant, only Boilers no. 3 and 4 are fitted with flue gas desulfurization (FGD) equipment.

FLUE GAS DESULFURIZATION SYSTEM

FGD Process Description. Initially, the FGD systems on both boilers were designed to operate by injection of dry limestone into the furnace, followed by tail gas scrubbing of both the SO₂ and the furnace-calcined limestone (as well as the fly ash)

from the flue gas stream. Because of tube pluggage in Boiler 4, attributed to limestone injection, this mode of operation was modified. Ground limestone is no longer injected into the furnace but instead is introduced into the flue gas near the gas inlet to the tail scrubber tower. The FGD system on Boiler 3 continued to operate as originally designed. Both systems, however, have undergone minor modifications to overcome such difficulties as build-up of sediment and pluggage of various components.

The FGD system on each boiler consists of two identical modules, each capable of treating 250,000 acfm of flue gas at 300°F. Bypassing of the modules is possible through a system of ducts and dampers around each module.

Details of a typical module are shown in Figures 12 and 13. Each scrubber module is 18 feet wide, 26 feet long, and 56 feet high. The lower 16 feet of the module is the reaction tank, in which the materials have time to complete their chemical reactions to remove sulfur dioxide. Two mixers are installed in the tank to provide thorough mixing and to keep the reactants and the fly ash in suspension, one on the lower back wall and the other on the upper front wall.

A sloping, perforated plate near the top of the reaction tanks acts as a strainer to catch large solid objects, such as scale, mud, or marbles, and funnels them to the mouth of a jaw crusher. Periodic operation of this equipment reduces these pieces to fine particles, which are discharged in a slurry to the fly ash pond.

As the flue gas enters the absorber tower above the liquid level in the reaction tank, it comes in contact with slurry, which is sprayed through 81 nozzles on 18 headers located under a marble bed. The marble bed consists of a 3- to 4-inch thickness of 3/4-inch-diameter glass marbles, retained on a perforated plate. Quenched flue gas and the sprayed slurry bubble through

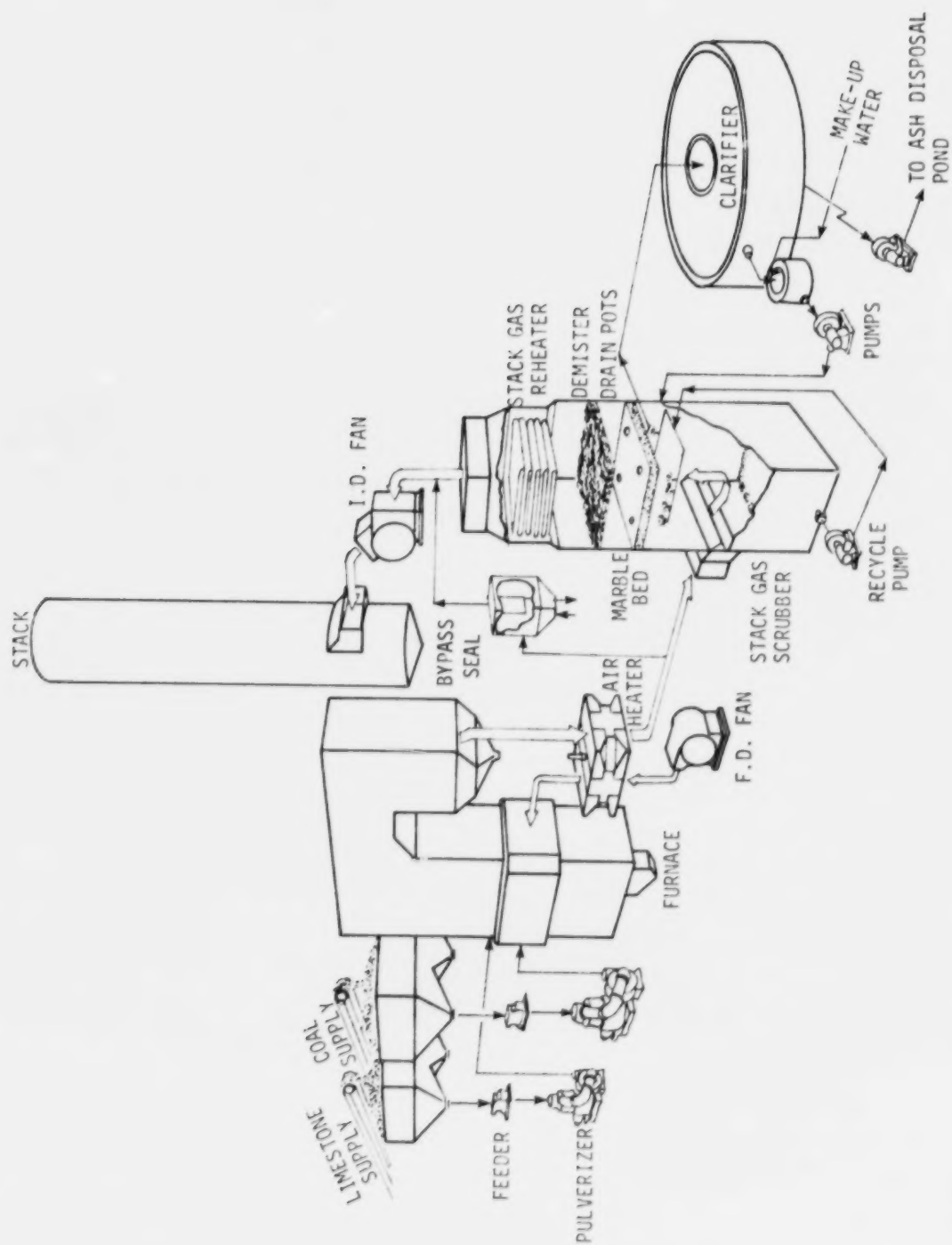


Figure 13. Sketch of a dry limestone furnace injection system with tail end scrubbing.

the marble bed, giving the marbles a spinning motion. The purpose of the moving marbles is to provide a large amount of surface area within a relatively small volume and thus to facilitate contact of the reactants. The liquid level above the marble bed is 10 to 13 inches high. The liquid overflow from the bed is discharged through 48 drain pots. The scrubbed flue gases flow upward 10 feet to a double row of Chevron separators designed to remove water vapors carried by the flue gases. The gases then pass around horizontal fin-tubed heaters that reheat the gases to about 175°F before they are discharged to the stack by the induced-draft fans. Gases from all modules on Boilers 3 and 4 are discharged to the atmosphere through a common stack 200 feet high.

Spent liquor from the reaction tank of each module is discharged to a 115-foot-diameter clarifier tank. The overflow from this tank flows to a clear well tank, where make-up water from the Missouri River is added. The discharge from this tank satisfies the water requirements of the FGD system. Distribution of the water is shown in Figure 12.

Sludge in the underflow from the clarifier tank is untreated. The thickened sludge is pumped to a 160-acre pond, which is also used for disposal of fly ash from the other boilers.

Inlet and outlet dampers are provided on each scrubber module to route the gases either through or around the module. To ensure passage of gases through the scrubber when it is in operation, a large U-shaped bypass seal is filled with water to positively close the bypass.

The demister wash system uses a set of eight water lances that turn on automatically to wash mud from the demisters whenever the dampers close. The lances are located four above and four below the demister vanes.

The reheater tubes are cleaned by two steam lances placed below the finned tubes. The heating medium is hot water from

the suction of the boiler feed pumps, delivered at 150 psig and 325°F.

FGD System Performance. Modifications recently completed by Combustion Engineering and Kansas City Power and Light Company apparently have significantly improved operation of the FGD systems. Module 3B on Boiler 3 recently operated for 550 hours with an availability factor of over 75 percent. Continuous monitoring equipment showed sulfur removal efficiency in the range of 85 to 95 percent. Particulate removal efficiency was 98 to 99 percent.

As a result of the recent modifications, the overall availability factor for the two modules on Boiler 3 has increased from about 30 percent to 70 percent. Availability of the FGD system for Boiler 4 has lagged because of the many problems attributed to dry limestone injection. Apart from those entailed in switching of Boiler 4 from furnace injection to tail gas injection, the major problems encountered with the two FGD systems at the Hawthorn Power Station have been identical. Analysis of the problems encountered during and since starting, reveals that nearly all were due to mechanical design rather than process chemistry.

1. Sediment Buildup: The reaction tank of each module is equipped with two mixers to keep the 10 percent solids in suspension. The original 15-hp mixers did not do this job properly, since buildup of hard mud, 6 to 8 feet high, developed in the corners of the reaction tank. Replacement of the original mixers with 25-hp units gave no significant improvement. The most recent modification aimed at eliminating sediment buildup consists of rounding off the bottom corners of the tanks by welding triangular steel plates, as shown in Figure 14. Further, a new make-up water piping system was installed, with four 1-inch nozzles located on the walls about 6 feet from the base of the tank and oriented to promote circulation and prevent settling of solids.

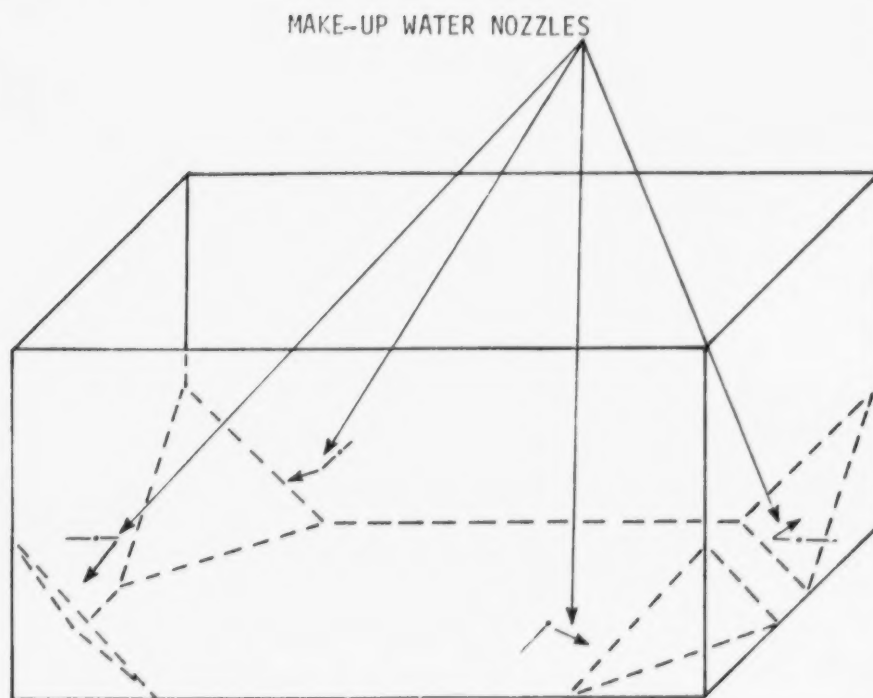


Figure 14. Sketch of the reaction tank showing the rounding off of the tank's corners and the installation of make-up water, sediment flushing system.

2. Pluggage of the Marble Bed: In the initial construction, the drain pots above the marble bed drained into horizontal headers that penetrated the module walls and emptied into the reaction tank. These horizontal headers plugged soon after the system was placed in operation. The headers were removed, and sloping headers were placed inside the reaction tank. Although these did not plug as fast, pluggage was still a problem. An additional problem occurred frequently when drain pots lifted off the marble bed allowed marbles to fall into the reaction tank. Both of these problems have been solved by installation of new stainless steel drain pots with expanded metal covers; a 3-foot rubber sock allows each drain pot to drain directly into the reaction tank. The pots are held in place by "J" bolts that attach to the marble bed. No pluggage of the new drain pots has occurred.

Pluggage in the marble bed has been alleviated also by an operational change. Increasing the liquid-to-gas ratio from 21 to 26 gallons per thousand cfm has increased the flow of liquid sufficiently to ensure that all parts of the marble bed are washed by at least two nozzles.

3. Excessive Wear of Spray Nozzles: Orifices of the spray nozzles were severely worn by abrasive action of the fly ash. The initial nozzles lasted only a few days; replacement was required because the enlarged orifices caused buildup of mud in the marble bed, overloading of the recycle pump, and clogging of the demister. Although KCPL installed a series of nozzles of different materials from different vendors, the longest life obtained from any of those was about 3 weeks. Since these nozzles are very expensive, KCPL designed a nozzle that could be made in the shop from pipe parts. These nozzles also lasted only about 3 weeks, but their cost was only 5 to 10 percent of the cost of vendors nozzles. Recently new ceramic nozzles were installed, and after 550 hours of operation they exhibited no signs of wear. Although this is not an exceptional service

time for nozzles, it is by far the best performance to date, since all other nozzles were completely worn in an equivalent amount of time.

4. Demister Pluggage: During sustained operations the scrubber modules were taken out of service every 3 days to allow the hosing of mud from the demisters. Pluggage has been controlled by adding retractable water lance blowers under the demister and by moving the rotary water lance blowers from below the demister to between the rows of demister vanes. With these modifications the demister wash system now performs satisfactorily.

5. Bypass Seal Pluggage: During periods when the flue gas is bypassed around the module, the drained-out water seal acts as a mechanical dust collector, accumulating fly ash. When the module is returned to operation and water is turned on, the fly ash in the water seal turns into mud, hampering operation of the bypass seal. Pluggage in this seal was eliminated by changing the flushing sequence within the seal.

REID GARDNER POWER STATION

Power Generating Facilities

The Reid Gardner Power Station of the Nevada Power Company is located near Moapa, Nevada, about 50 miles north of Las Vegas. The station has two electric power generating units (Reid Gardner 1 and 2), each rated at 126 MW. The units are identical in design, and both operate under base load conditions. The coal-fired boilers were manufactured by Foster Wheeler. Reid Gardner 1 was placed in service in 1965, and Reid Gardner 2, in 1968. Average gross heating value of the coal burned is 12,450 BTU/lb; average ash and sulfur contents are 9 and 0.6 percent, respectively.

Reid Gardner 3, with design capacity of 125 MW, is under construction, with start-up scheduled for the spring of 1975. Reid Gardner 4, also a 125-MW unit, is in the planning stage, with operation expected in 1978.

FLUE GAS DESULFURIZATION SYSTEM

The sodium carbonate base flue gas desulfurization (FGD) systems on Reid Gardner 1 and 2 were manufactured by Combustion Equipment Associates and placed in operation during March and April 1974. The design of these systems was based on data gathered from operation of an 8000-acfm pilot plant, which operated at the Reid Gardner Power Station during 1971 and 1972.

FGD Process Description. The FGD system on each boiler consists of a single module, designed to handle 473,000 acfm of gas at 350°F. The module is made up of a twin variable-throat venturi scrubber followed by a single-stage SO₂ scrubber tower. The FGD modules on both boilers share a common trona (sodium carbonate ore) storage and beneficiation plant. Bypassing of each

module is possible through a guillotine-type bypass valve. As shown in Figure 15, the hot flue gas from the boiler passed first through an existing mechanical collector (multicyclone), where about 75 percent of the fly ash is removed. Pressure is then boosted by an induced-draft fan before the stream splits as it enters the twin variable-throat venturi scrubber. The hot flue gas is quenched by a shower of circulated sodium-base liquor, sprayed from tangential nozzles located on the walls of the venturi. The gas-liquid mixture then passes through the venturi throat and exits into the sump, where the reduction in velocity causes the liquor droplets to fall and separate from the gas stream. A liquid-to-gas ratio of 10 gallons/ 1000 acfm and a pressure drop of about 15 inches of water are maintained in the venturi for efficient removal of SO_2 and particulates. The scrubbed gas then enters the SO_2 scrubber tower tangentially. The centrifugal force created causes the finer droplets to coalesce near the tower's wall and separate from the gas stream. The scrubbed gas, which is now essentially free of fly ash and of most of its sulfur dioxide content, rises through the tower and bubbles through the 3/8-inch-diameter holes of a single sieve tray. The tray is flooded with clear water from the ash pond at a rate equivalent to 1 gpm per 1000 acfm of gas and is intermittently washed with fresh water from below. More SO_2 is absorbed in the clear water tray and the pH of the liquid effluent drops to about 5. The cleaned gas continues up the tower and passes through the radial-vane-type mist eliminator, where the liquid droplets carried over with the gas are trapped and removed.

The gas is reheated by direct contact with hot air to provide buoyancy and to prevent condensation of the gas on duct walls or in the stack. Ambient air is drawn in by a separate fan and steam-heated in a shell and tube heat exchanger before it intermixes with the scrubbed wet gas in the outlet duct of the module. The combined gas stream, which is now warmer, enters the 200-foot-tall stack and discharged to the atmosphere.

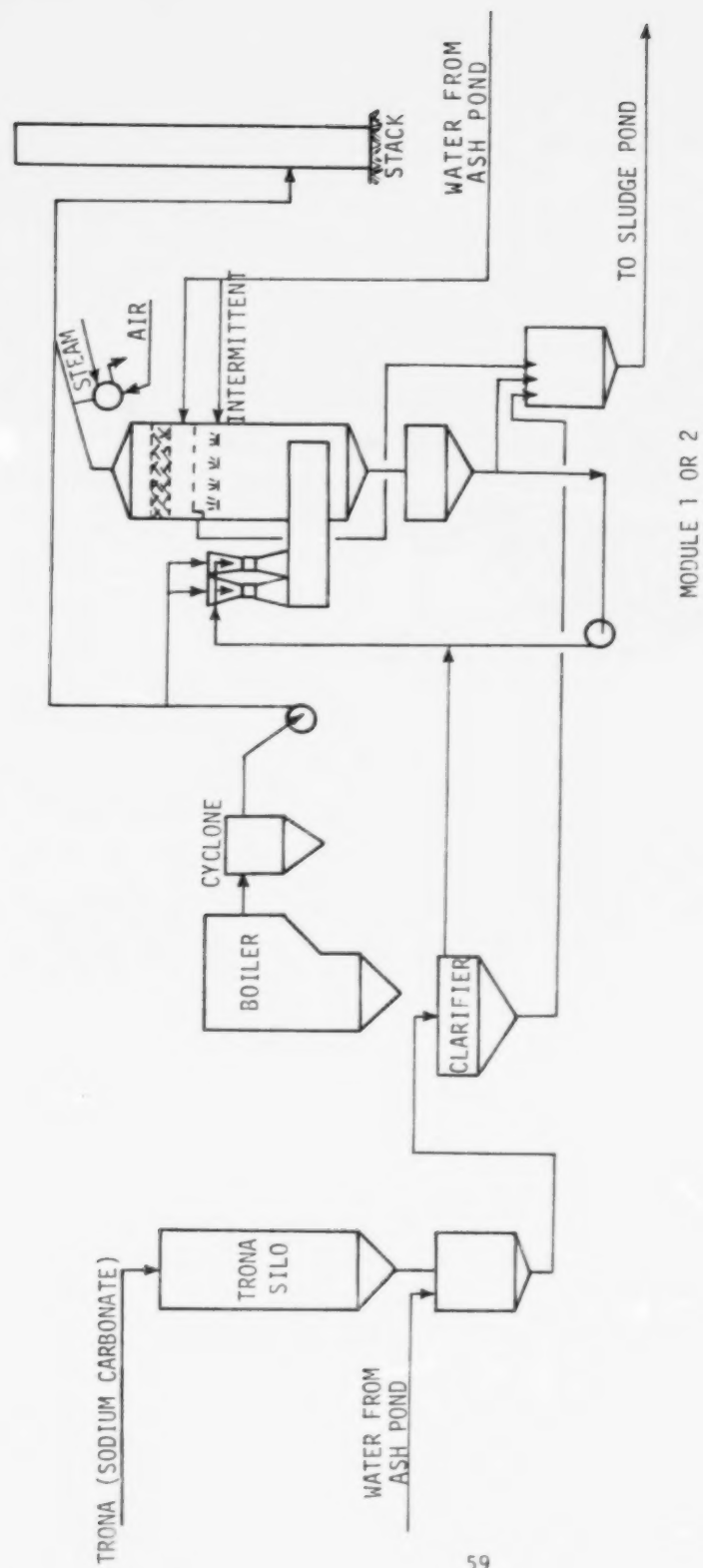


Figure 15. Simplified flow diagram of the FGD system on each of Reid Gardner 1 and 2 electric power generating units.

The SO₂ scrubbing reagent in the FGD systems is trona, an impure form of sodium carbonate. It is mined by Morrison and Weatherly Chemical Products Company from Owens Lake in California and shipped to the Reid Gardner Power Station as a powder. The trona is transferred pneumatically to a tall silo having a 48-hour-supply storage capacity. It is then transferred to the day bin from which it is conveyed by screw conveyor to a slurry tank and mixed with fresh water to dissolve its sodium carbonate content. The slurry, containing such impurities as sand and sodium chloride, is pumped into a clarifier, where the insoluble impurities settle out. The clarified sodium carbonate liquor is then pumped to the modules recirculation tanks. The effluent from the FGD system is made up of three sources; the slip stream of spent liquor discharged from the recirculation tank, the acidic effluent from the clear water tray, and the alkaline clarifier underflow which serves to neutralize the pH of the combined liquor before it is discharged to the sludge settling ponds. The spent liquor is pumped to a small 6-acre pond, from which overflow can flow to a 45-acre pond. Because operation of the FGD system has been limited and the evaporation rate from the 6-acre pond is high, no liquor has yet trickled to the larger pond. The water system operates on an open loop, with no liquor recycled to the modules from the ponds.

FGD System Performance. Since their start-up in early 1974, both modules at the Reid Gardner Power Station have performed satisfactorily. Because of difficulties in obtaining sufficient quantities of trona, however, operation of the modules has been subjected to frequent interruptions and the flue gas has been bypassed to conserve trona. The longest continuous operation was on Module 2, for 18 days. The FGD system has never had a forced shutdown due to problems originating in the equipment, and thus can be considered as essentially 100 percent available, with operations interrupted only by the trona shortage. Since start-up, each module has operated for 900 hours.

Since the major problem preventing continuous operation of the FGD modules has been and still is the inadequate supply of trona, the Nevada Power Company has been experimenting with the use of sodium carbonate brine. One tank truck shipment of brine from Green River, Wyoming, has been delivered and will soon be tested on Reid Gardner 2 unit. Larger quantities cannot be received, since the brine freezes at ambient temperature and the plant lacks the proper facilities for storage.

The system experienced the usual minor start-up problems, such as rubber liners on recirculation lines becoming loose as a result of poor fabrication, freeze-up of instruments, plugging in bypass lines of some control valves, and leakage in the mechanical seals of the recirculation pumps. Further freeze-up problems were experienced with the louver type bypass dampers, which were later replaced by a combination of louver/guillotine-type dampers.

As expected, this soluble-base FGD system has not experienced such problems as equipment scaling, demister plugging, erosion, corrosion, which are normally associated with lime/limestone FGD systems.

INFORMATION SOURCES

Much of the information reported here concerning FGD systems operation was obtained recently by visits to plant installations and discussions with plant personnel:

Hawthorn Power Station	June 6, 1974
Will County Power Station	June 20, 1974
Lawrence Power Station	August 13, 1974
Reid Gardner	October 8, 1974

In addition, the following reports are cited for further references:

- 1) Palermo, Joseph V. "Operating Experience on SO₂ Scrubbers, Hawthorn Units 3 and 4" Report to Missouri Valley Electric Association at Kansas City, Missouri on April 18, 1974.
- 2) Gifford, D.C. "Will County Unit 1, Limestone Wet Scrubbers, Description and Operating Experiences: November 30, 1973.
- 3) Gifford, D.C. "Will County Unit 1, Limestone Wet Scrubber, Waste Sludge Disposal" Paper presented at the Electrical world Symposium in Chicago, Illinois on October 30-31, 1973.
- 4) Green, Kelly. "Operating Experience with Limestone Injection-Wet Scrubbings Stack Gas Treatment" Paper presented at the Frontiers of Power Technology Conference at Stillwater, Oklahoma on October 10-11, 1973.

60

TABLE 1
SUMMARY OF FGD SYSTEMS

FGD STATUS REPORT 10/74

I.D. NUMBER AND COMPANY NAME	UNIT NAME	LOCATION	START UP DATE	STATUS
1 ALABAMA ELECTRIC COOP	BISSIE NO 2	JACKSON ALABAMA	1-79	6
2 ALABAMA ELECTRIC COOP	BISSIE NO 3	JACKSON ALABAMA	1-79	6
3 ARIZONA ELECTRIC POWER	APACHE NO 2	CUCUMISE ARIZONA	6-79	4
4 ARIZONA ELECTRIC POWER	APACHE NO 3	CUCUMISE ARIZONA	6-79	4
5 ARIZONA PUBLIC SERVICE	CHOLLA NO 1	JOSEPH CITY ARIZONA	10-73	1
6 ARIZONA PUBLIC SERVICE	CHOLLA NO 2	JOSEPH CITY ARIZONA	6-77	3
7 ARIZONA PUBLIC SERVICE	CHOLLA NO 3	JOSEPH CITY ARIZONA	6-78	3
8 ARIZONA PUBLIC SERVICE	CHOLLA NO 4	JOSEPH CITY ARIZONA	6-79	6
9 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 1	FARMINGTON NEW MEXICO	10-75	6
10 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 2	FARMINGTON NEW MEXICO	12-76	6
11 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 3	FARMINGTON NEW MEXICO	3-77	6
12 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 4	FARMINGTON NEW MEXICO	4-77	6
13 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 5	FARMINGTON NEW MEXICO	6-77	6
14 BASIN ELECTRIC	MISSOURI BASIN NO 1	WHEATLAND WYOMING	6-79	6
15 BASIN ELECTRIC	MISSOURI BASIN NO 2	WHEATLAND WYOMING	6-80	6
16 BASIN ELECTRIC	MISSOURI BASIN NO 3	WHEATLAND WYOMING	6-83	6
17 BOSTON Edison	MYSTIC NO 8	EVERETT MASSACHUSETTS	4-72	1
18 CENTRAL ILLINOIS LIGHT CO.	DUCK CREEK NO.1	CANTON ILLINOIS	3-76	2
19 CENTRAL ILLINOIS PUBLIC SERV	NEWTON NO.1	ALTON ILLINOIS	3-77	3
20 CENTRAL ILLINOIS GAS AND ELECTRIC	EAST RENO NO 1	RABBIT HASH KENTUCKY	6-79	6
21 CENTRAL ILLINOIS GAS AND ELECTRIC	MIAMI FORT NO 6	NORTH BEND OHIO	1-77	5
22 CITY OF NEW YORK	KEY WEST POWER PLANT	KEY WEST FLORIDA	10-72	1
23 COLORADO UTL ELECTRIC	CRAIG STATION NO.1	CRAIG COLORADO	6-78	6
24 COLORADO UTL ELECTRIC	CRAIG STATION NO.2	CRAIG COLORADO	6-79	6
25 COLORADO UTL ELECTRIC	HAYDEN NO 1	HAYDEN COLORADO	6-78	6
26 COLORADO UTL ELECTRIC	HAYDEN NO 2	HAYDEN COLORADO	6-78	6
27 COLORADO & MICHIGAN	CONESVILLE NO 3	CONESVILLE OHIO	6-78	3
28 COLORADO & MICHIGAN	CONESVILLE NO 6	CONESVILLE OHIO	1-78	3
29 COMINGHEALTH Edison	WILL COUNTY NO 1	ROSEMOUNT ILLINOIS	2-72	1
30 CALIFORNIA POWER COOP	ALMA STATION	ALMA WISCONSIN	6-71	1
31 CALIFORNIA POWER COOP	ST. CLAIR NO 6	ELLE RIVER MICHIGAN	11-74	2
32 CALIFORNIA POWER COOP	CLARK	ELKAPA PENNSYLVANIA	7-73	2
33 CALIFORNIA POWER COOP	PHILLIPS	SOUTH HEIGHT PENNSYLVANIA	3-74	1
34 CALIFORNIA POWER COOP	CHEVROLET PARKA 1 2 3 & 4	PARKA OHIO	4-77	5
35 CALIFORNIA POWER COOP	MOORE CITY NO.3	MOORE CITY PENNSYLVANIA	10-72	1
36 CALIFORNIA POWER COOP	WOOD RIVER NO 4	EAST ALTON ILLINOIS	4-77	5
37 INDIANAPOLIS POWER AND LIGHT	PETERSBURG NO 3	PETERSBURG INDIANA	11-72	1
38 KANSAS CITY POWER & LIGHT	WATHORN NO 4	KANSAS CITY MISSOURI	6-72	1
39 KANSAS CITY POWER & LIGHT	WATHORN NO 1	KANSAS CITY MISSOURI	6-73	1
40 KANSAS CITY POWER & LIGHT	LA CYGNE NO 1	LA CYGNE KANSAS	6-78	3
41 KANSAS CITY POWER & LIGHT	JEFFERY NO. 1	ST MARY KANSAS	6-79	5
42 KANSAS CITY POWER & LIGHT	JEFFERY NO. 2	ST MARY KANSAS	6-79	5

1. OPERATIONAL UNITS
2. UNITS UNDER CONSTRUCTION
3. PLANNED - CONTRACT AWARDED
4. PLANNED - LETTER OF INTENT SIGNED
5. PLANNED - REQUESTING/EVALUATING BIOS
6. CONSIDERING ONLY FGD SYSTEMS

90

TABLE 1
SUMMARY OF FGD SYSTEMS

FGD STATUS REPORT 10/74

1. U. NUMBER AND COMPANY NAME	UNIT NAME	LOCATION	START UP DATE	STATUS
43 KANSAS POWER & LIGHT	LAWRENCE NO 4	LAWRENCE KANSAS	12-68	1
44 KANSAS POWER & LIGHT	LAWRENCE NO 5	LAWRENCE KANSAS	11-71	1
45 KENTUCKY UTILITIES	GREEN RIVER 1 2 & 3	SOUTH CAMLTON KENTUCKY	4-75	2
46 LOUISVILLE GAS & ELECTRIC	CAME RUN NO 1	LOUISVILLE KENTUCKY	6-60	6
47 LOUISVILLE GAS & ELECTRIC	CAME RUN NO 2	LOUISVILLE KENTUCKY	6-60	6
48 LOUISVILLE GAS & ELECTRIC	CAME RUN NO 3	LOUISVILLE KENTUCKY	6-60	6
49 LOUISVILLE GAS & ELECTRIC	CAME RUN NO 4	LOUISVILLE KENTUCKY	6-75	3
50 LOUISVILLE GAS & ELECTRIC	CAME RUN NO 5	LOUISVILLE KENTUCKY	6-76	6
51 LOUISVILLE GAS & ELECTRIC	CAME RUN NO 6	LOUISVILLE KENTUCKY	6-77	6
52 LOUISVILLE GAS & ELECTRIC	MILL CREEK NO 1	LOUISVILLE KENTUCKY	6-79	6
53 LOUISVILLE GAS & ELECTRIC	MILL CREEK NO 2	LOUISVILLE KENTUCKY	6-78	6
54 LOUISVILLE GAS & ELECTRIC	MILL CREEK NO 3	LOUISVILLE KENTUCKY	6-77	3
55 LOUISVILLE GAS & ELECTRIC	MILL CREEK NO 4	LOUISVILLE KENTUCKY	6-79	6
56 LOUISVILLE GAS & ELECTRIC	PAIDOTS RUN NO 6	LOUISVILLE KENTUCKY	4-73	1
57 Louisiana Power	CULSTRIP NO 1	CULSTRIP MONTANA	5-75	2
58 Montana Power	CULSTRIP NO 2	CULSTRIP MONTANA	5-76	2
59 Montana Power	CULSTRIP NO 3	CULSTRIP MONTANA	7-78	6
60 Montana Power	CULSTRIP NO 4	CULSTRIP MONTANA	0-79	6
61 Nevada Power	KLIO GARDNER NO 1	MAOPA NEVADA	12-73	1
62 Nevada Power	KLIO GARDNER NO 2	MAOPA NEVADA	12-73	1
63 Nevada Power	KLIO GARDNER NO 3	MAOPA NEVADA	6-75	2
64 Nevada Power	KLIO GARDNER NO 4	MAOPA NEVADA	6-76	4
65 NEW JERSEY LFC SYSTEM	BRAYTON POINT NO.3	SOMERSET MASSACHUSETTS	0-0	6
66 Northern Indiana Pub Service	DAILY NO.7	CHESTERTON INDIANA	0-0	6
67 Northern Indiana Pub Service	DAILY NO.8	CHESTERTON INDIANA	0-0	6
68 Northern Indiana Pub Service	PITCHER NO 11	GARY INDIANA	6-75	2
69 Northern Indiana Pub Service	SHERBORN NO 1	HECKER MINNESOTA	5-78	2
70 Pennsylvania Power Co.	SHERBORN NO 2	HECKER MINNESOTA	5-77	2
71 Pennsylvania Power Co.	BRUCE MANSFIELD NO. 1	SHIPPINGPORT PENNSYLVANIA	6-75	2
72 Pennsylvania Power Co.	BRUCE MANSFIELD NO. 2	SHIPPINGPORT PENNSYLVANIA	6-75	2
73 Pennsylvania Power Co.	BRUCE MANSFIELD NO. 3	SHIPPINGPORT PENNSYLVANIA	8-78	6
74 Philadelphia Electric	EDDYSTONE NO 1	EDDYSTONE PENNSYLVANIA	12-74	2
75 Potomac Electric & Power	DICKERSON NO 3	DICKERSON MARYLAND	9-73	1
76 Public Service of Indiana	GLOSS NO. 2	PRINCETON INDIANA	5-75	2
77 Public Service of New Mexico	SAT JUAN NO 1	WATERFLOW NEW MEXICO	10-77	3
78 Public Service of New Mexico	SAT JUAN NO 2	WATERFLOW NEW MEXICO	1-77	3
79 Public Service of New Mexico	SAT JUAN NO.3	WATERFLOW NEW MEXICO	5-78	6
80 Public Service of New Mexico	SAT JUAN NO.4	WATERFLOW NEW MEXICO	5-80	6
81 S. Carolina Pub Serv Authority	GLORGETOWN NO 2	GLORGETOWN SOUTH CAROLINA	5-77	5
82 Salt River Project	NAVAJO NO 1	PAGE ARIZONA	3-76	6
83 Salt River Project	NAVAJO NO 2	PAGE ARIZONA	10-76	6
84 Salt River Project	NAVAJO NO 3	PAGE ARIZONA	3-77	6

1. OPERATIONAL UNITS
2. UNITS UNDER CONSTRUCTION
3. PLANNED - CONTRACT AWARDED
4. PLANNED - LETTER OF INTENT SIGNED
5. PLANNED - REQUESTING/EVALUATING BIDS
6. CONSIDERING ONLY FGD SYSTEMS

90

TABLE 1
SUMMARY OF FGD SYSTEMS

FGD STATUS REPORT 10/74	I.D. NUMBER	COMPANY NAME	UNIT NAME	LOCATION	START UP DATE	STATUS
	85	SOUTHERN CALIFORNIA Edison	ROMNEY NO. 1	LAURELIN NEVADA	6-77	6
	86	SOUTHERN CALIFORNIA Edison	ROMNEY NO. 2	LAURELIN NEVADA	12-74	2
	87	SOUTHERN CALIFORNIA Edison	ROMNEY NO. 3	LAURELIN NEVADA	11-75	1
	88	SOUTHERN CALIFORNIA Edison	ROMNEY NO. 4	LAURELIN NEVADA	6-77	6
	89	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 1	HATTIESBURG MISSISSIPPI	6-79	5
	90	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 2	HATTIESBURG MISSISSIPPI	6-79	5
	91	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 3	HATTIESBURG MISSISSIPPI	6-79	5
	92	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 4	HATTIESBURG MISSISSIPPI	6-79	5
	93	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 5	HATTIESBURG MISSISSIPPI	6-79	5
	94	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 6	HATTIESBURG MISSISSIPPI	6-79	5
	95	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 7	HATTIESBURG MISSISSIPPI	6-79	5
	96	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 8	HATTIESBURG MISSISSIPPI	6-79	5
	97	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 9	HATTIESBURG MISSISSIPPI	6-79	5
	98	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 10	HATTIESBURG MISSISSIPPI	6-79	5
	99	SOUTHERN MISSISSIPPI PER COOP	HATTIESBURG NO. 11	HATTIESBURG MISSISSIPPI	6-79	5

1. OPERATIONAL UNIT 4. PLANNED - LETTER OF INTENT SIGNED
 2. UNIT UNDER CONSTRUCTION 5. PLANNED - REQUESTING/EVALUATING BIDS
 3. PLANNED - CONTRACT AWARD 6. CONSIDERING ONLY FGD SYSTEMS

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
ARIZONA ELECTRIC COOP	10. NUMBER 1	INSTALLATION OF A FLUE GAS DESULFURIZATION SYSTEM IS BEING CONSIDERED.
	APACHE NO 2	IF A SYSTEM IS INSTALLED, IT WILL PROBABLY BE THE NET LIMESTONE SCRUBBING PROCESS.
	205 NO. 2 - NEW	
	COAL 0.9 - 1.5 PERCENT SULFUR	
ARIZONA ELECTRIC COOP	NOT SELECTED	
	LIMESTONE SCRUBBING	
	STARTUP 1/79	
	205 NO. 3 - NEW	
ARIZONA ELECTRIC COOP	10. NUMBER 2	INSTALLATION OF A FLUE GAS DESULFURIZATION SYSTEM IS BEING CONSIDERED.
	APACHE NO 3	IF A SYSTEM IS INSTALLED, IT WILL PROBABLY BE THE NET LIMESTONE SCRUBBING PROCESS.
	205 NO. 3 - NEW	
	COAL 0.8 - 1.5 PERCENT SULFUR	
ARIZONA ELECTRIC COOP	NOT SELECTED	
	LIMESTONE SCRUBBING	
	STARTUP 1/79	
	205 NO. 3 - NEW	
ARIZONA ELECTRIC COOP	10. NUMBER 3	LETTER OF INTENT TO INSTALL AN FGD SYSTEM WAS ISSUED TO RESEARCH COTTRELL.
	APACHE NO 3	
	205 NO. 3 - NEW	
	COAL 0.5 - 0.8 PERCENT SULFUR	
ARIZONA ELECTRIC COOP	NOT SELECTED	
	LIMESTONE SCRUBBING	
	STARTUP 6/79	
	205 NO. 3 - NEW	
ARIZONA ELECTRIC COOP	10. NUMBER 4	LETTER OF INTENT TO INSTALL AN FGD SYSTEM WAS ISSUED TO RESEARCH COTTRELL.
	APACHE NO 4	
	205 NO. 4 - NEW	
	COAL 0.5 - 0.8 PERCENT SULFUR	
ARIZONA ELECTRIC COOP	NOT SELECTED	
	LIMESTONE SCRUBBING	
	STARTUP 6/79	
	205 NO. 4 - NEW	
ARIZONA ELECTRIC COOP	10. NUMBER 5	BOTH MODULES WERE IN SERVICE DURING THE PAST 30 DAYS, EXCEPT FOR ONE DAY PER MODULE TO CORRECT A VIBRATION PROBLEM IN THE REHEATER UNITS. THE INSTALLATION OF GAFFLE PLATES IN THE INLET DUCT TO EACH REHEATER TO DAMPEN THE VIBRATION IS CONSIDERED A PARTIAL SUCCESS BY APS. FGD SYSTEM AVAILABLE WAS 93 PERCENT.
	APACHE NO 5	
	205 NO. 5 - NEW	
	COAL 0.9 - 1.5 PERCENT SULFUR	
ARIZONA ELECTRIC COOP	NOT SELECTED	
	LIMESTONE SCRUBBING	
	STARTUP 10/73	
	205 NO. 5 - NEW	

Table 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
ARIZONA PUBLIC SERVICE	CHOLLA NO 2	CONTRACT TO INSTALL FGD SYSTEM ON CHOLLA NO 3 HAS BEEN AWARDED TO RESEARCH CONTRELL.
CHOLLA NO 2	250 MW - NEW	
COAL	0.44 PERCENT SULFUR	
RESEARCH CONTRELL		
LIVESTONE SCRUBBING		
STARTUP	6/77	
ARIZONA PUBLIC SERVICE	CHOLLA NO 3	ARIZONA PUBLIC SERVICE IS OUT FOR BIDS AND SOME PROPOSALS HAVE BEEN RECEIVED AND ARE BEING EVALUATED.
CHOLLA NO 3	250 MW - CA	
COAL	0.44 PERCENT SULFUR	
NOT SELECTED		
LIVESTONE SCRUBBING		
STARTUP	6/78	
ARIZONA PUBLIC SERVICE	CHOLLA NO 4	PLANS FOR SO2 CONTROL SYSTEM ARE IN THE EARLY STAGE. APS IS CURRENTLY WORKING ON THE DESIGN OF UNITS 2 AND 3 AND HAS NOT FINNED UP PLANS FOR UNIT 4.
CHOLLA NO 4	512 MW - NEW	
COAL		
NOT SELECTED		
LIVESTONE SCRUBBING		
STARTUP	6/79	
ARIZONA PUBLIC SERVICE	CHOLLA NO 5	EXISTING SCRUBBERS REMOVE ABOUT 30 PERCENT OF THE SO2. THE VENTURIS WILL BE UPGRADED TO COMPLY WITH APPLICABLE SO2 REGULATIONS.
CHOLLA NO 5	175 MW - RETROFIT	
COAL	0.8 PERCENT SULFUR	
CHEMICO		
LIFE SCRUBBING		
STARTUP	12/76	
ARIZONA PUBLIC SERVICE	CHOLLA NO 6	EXISTING SCRUBBERS REMOVE ABOUT 30 PERCENT OF THE SO2. THE VENTURIS WILL BE UPGRADED TO COMPLY WITH APPLICABLE SO2 REGULATIONS.
CHOLLA NO 6	175 MW - RETROFIT	
COAL	0.8 PERCENT SULFUR	
CHEMICO		
NOT SELECTED		
STARTUP	12/76	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
ARIZONA PUBLIC SERVICE	10. NUMBER 11	EXISTING SCRUBBERS REMOVE ABOUT 30 PERCENT OF THE SO ₂ . THE VENTURIS WILL BE UPGRADED TO COMPLY WITH APPLICABLE SO ₂ REGULATIONS.
FOUR CORNERS NO 3	229 PA - RETROFIT	
COAL	0.0 PERCENT SULFUR	
CHEMICO	LIME SCRUBBING	
STARTUP	5/77	
ARIZONA PUBLIC SERVICE	12. NUMBER 12	LSP'S ARE USED FOR PARTICULATE EMISSION CONTROL. FGD SYSTEM IS STILL IN THE PLANNING STAGE.
FOUR CORNERS NO 4	600 PA - RETROFIT	
COAL	0.0 PERCENT SULFUR	
CHEMICO	LIME SCRUBBING	
STARTUP	4/77	
ARIZONA PUBLIC SERVICE	13. NUMBER 13	LSP'S ARE USED FOR PARTICULATE EMISSION CONTROL. FGD SYSTEM IS STILL IN THE PLANNING STAGE.
FOUR CORNERS NO 5	600 PA - RETROFIT	
COAL	0.0 PERCENT SULFUR	
NOT SELECTED	LIME/LIMESTONE SCRUBBING	
STARTUP	6/77	
MISSOURI BASIN NO 1	14. NUMBER 14	THE COMPANY IS EVALUATING ALTERNATIVE PROCESSES FOR SO ₂ EMISSION CONTROL. LIMESTONE SCRUBBING IS THE MOST LIKELY PROCESS TO BE SELECTED. THE FINAL DECISION WILL BE MADE IN A FEW MONTHS.
550 PA - NEW		
COAL	NOT SELECTED	
LIMESTONE SCRUBBING		
STARTUP	6/79	
MISSOURI BASIN NO 2	15. NUMBER 15	THE COMPANY IS EVALUATING ALTERNATIVE PROCESSES FOR SO ₂ EMISSION CONTROL. LIMESTONE SCRUBBING IS THE MOST LIKELY PROCESS TO BE SELECTED. THE FINAL DECISION WILL BE MADE IN A FEW MONTHS.
550 PA - NEW		
COAL	NOT SELECTED	
LIMESTONE SCRUBBING		
STARTUP	6/80	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10774

COMPANY	POWER STATION	CURRENT MONTH
1.0. NUMBER 16 WASIN ELECTRIC MISSOURI BASIN NO 5 550 MW - NEW	COAL NOT SELECTED LIMESTONE SCRUBBING STARTUP 6/83	THE COMPANY IS EVALUATING ALTERNATIVE PROCESSES FOR SO ₂ EMISSION CONTROL. LIMESTONE SCRUBBING IS THE MOST LIKELY PROCESS TO BE SELECTED. THE FINAL DECISION WILL BE MADE IN A FEW MONTHS.
1.0. NUMBER 17 GASTIG LUISON MYSTIC NO 6 150 MW - RETROFIT OIL 2.5 PERCENT SULFUR	COAL NOT SELECTED LIMESTONE SCRUBBING STARTUP 4/72	FGD SYSTEM IS DOWN WITH NO IMMEDIATE PLANS FOR RESTARTING BECAUSE EPA FUNDING HAS EXPIRED AND THE CALCINING PLANT IS NOW BEING USED BY ANOTHER COMPANY. THE UTILITY IS EVALUATING THE DATA COLLECTED DURING THE DEMONSTRATION PERIOD TO DETERMINE THE COURSE OF FUTURE ACTION.
1.0. NUMBER 18 CENTRAL ILLINOIS LIGHT CO. DUCK CREEK NO.1 100 MW - NEW COAL 2.5-3.0 PERCENT SULFUR KILEY STOCKPILING/ENGINEERING LIMESTONE SCRUBBING STARTUP 3/76	COAL NOT SELECTED LIMESTONE SCRUBBING STARTUP 4/72	THE UNIT IS NOW UNDER CONSTRUCTION AND IS SCHEDULED FOR COMPLETION IN MARCH 1976. STATION CAPACITY IS 400MW, BUT FLUE GAS FROM ONLY 100 MW WILL BE TREATED INITIALLY.
1.0. NUMBER 19 CENTRAL ILLINOIS PUBLIC SERV MEATON NO.1 600 MW - NEW COAL 2.0-3.0 PERCENT SULFUR NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 3/77	COAL NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 3/77	BIDS HAVE BEEN TAKEN ON LIME AND LIMESTONE SCRUBBERS AND ARE NOW BEING EVALUATED.
1.0. NUMBER 20 CINCINNATI GAS AND ELECTRIC EAST WEND NO.1 600 MW - NEW COAL 0.5-0.8 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 6/79	COAL NOT SELECTED NOT SELECTED STARTUP 6/79	THE COMPANY IS PRESENTLY OBTAINING CONSTRUCTION AND OPERATION PERMITS. THEY PLAN TO GO OUT FOR BIDS ON AN FGD SYSTEM.

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
1.0. NUMBER 21	CINCINNATI GAS AND ELECTRIC	
500 MA - NEW	MIAMI FORT WORTH	
COAL 0.2 PERCENT SULFUR		THE COMPANY IS OBTAINING BIDS FOR AN FGD SYSTEM.
NOT SELECTED		
LIMESTONE SCRUBBING		
STARTUP 1/77		
1.0. NUMBER 22	CITY OF KEY WEST	
KEY WEST POWER PLANT		PLANT HAS BEEN SHUT DOWN SINCE SEPT 19TH BECAUSE OF A BOILER 1 D FAN PROBLEM FAILURE. THEY HOPE TO BE OPERATIONAL THE WEEK OF OCTOBER 21ST. MECHANICAL FAILURES IN THE FGD SYSTEM WERE FIXED. FGD OPERATED ABOUT A WEEK AND A HALF BEFORE THE PRESENT BOILER SHUTDOWN. DURING SHUTDOWN THE EXHAUSTION COILS ON THE DUCTWORK TO THE SCRUBBER WERE REPLACED FOR THE 2ND TIME. SOME PLUGGING REMOVED AT SHUTDOWN.
37 MA - NEW		
COAL 2.5 PERCENT SULFUR		
NOT SELECTED		
LIMESTONE SCRUBBING		
STARTUP 10/72		
1.0. NUMBER 23	COLORADO UTE ELECTRIC	
CRAIG STATION NO. 1		AS WITH HAYUEM 1 & 2 BOTH LIMESTONE AND SODIUM CARBONATE SYSTEMS ARE BEING EVALUATED. USING THE ANTICIPATED 0.45 PERCENT SULFUR COAL AN FGD SYSTEM WITH APPROX. 70 PERCENT EFF. WILL BE REQUIRED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. PRESENT EVALUATION FAVORS A LIMESTONE SCRUBBER WITH SLUDGE DISPOSAL IN A MINE SITE.
450 MA - NEW		
COAL 0.45 PERCENT SULFUR		
NOT SELECTED		
LIMESTONE SCRUBBING		
STARTUP 5/72		
1.0. NUMBER 24	COLORADO UTE ELECTRIC	
CRAIG STATION NO. 2		AS WITH HAYUEM 1 & 2 BOTH LIMESTONE AND SODIUM CARBONATE SYSTEMS ARE BEING EVALUATED. USING THE ANTICIPATED 0.45 PERCENT SULFUR COAL AN FGD SYSTEM WITH APPROX. 70 PERCENT EFF. WILL BE REQUIRED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. PRESENT EVALUATION SEEMS TO FAVOR A LIMESTONE SCRUBBER WITH SLUDGE DISPOSAL IN A MINE SITE.
450 MA - NEW		
COAL 0.45 PERCENT SULFUR		
NOT SELECTED		
LIMESTONE SCRUBBING		
STARTUP 6/78		
1.0. NUMBER 25	COLORADO UTE ELECTRIC	
HAYUEM NO. 1		COLORADO UTE IS PROCEEDING WITH ENGINEERING EVALUATION OF BOTH LIMESTONE & SODIUM CARBONATE SYSTEMS. USING THE PRESENT 0.45 PERCENT SULFUR COAL AN FGD UNIT WITH APPROX. 70 PERCENT EFF. WOULD BE NEEDED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. IT IS PROBABLE THAT A LIMESTONE SCRUBBER WILL BE CHOSEN WITH DISPOSAL IN A MINE SITE.
202 MA - RETROFIT		
COAL 0.45 PERCENT SULFUR		
NOT SELECTED		
LIMESTONE SCRUBBING		
STARTUP 9/79		

TABLE 2
STATUS OF FGD SYSTEMS DURING 1977

CORPORATION	POWER STATION	CURRENT MONTH
1.0. NUMBER 28	COLORADO UTE ELECTRIC	COLORADO UTE IS PROCEEDING WITH ENGINEERING EVALUATION OF BOTH LIME- STONE & SODIUM CARBONATE SYSTEMS. USING THE PRESENT 0.45 PERCENT SULFUR COAL AN FGD UNIT WITH APPROX. 70 PERCENT EFF. WOULD BE NEEDED TO MEET THE 1978 COLORADO SO2 EMISSION REGS OF 150PPM. IT IS PROBABLE THAT A LIMESTONE SCRUBBER WILL BE CHOSEN WITH DISPOSAL IN A MINE SITE.
250 MW - RETROFIT	COAL 0.45 PERCENT SULFUR	
NOT SELECTED	LIMESTONE SCRUBBING	
STARTUP 6/78		
1.0. NUMBER 27	COLUMBUS & SOUTHERN OHIO	U.O.P. IS PRESENTLY IN THE ENGINEERING DESIGN AND PROCUREMENT STAGE. CONSTRUCTION IS SCHEDULED TO START MARCH 1975.
COLUMBIAVILLE NO 5	375 MW - NEW	
COAL	UNIVERSAL OIL PRODUCTS	
LIME SCRUBBING		
STARTUP 1/78		
1.0. NUMBER 26	COLUMBUS & SOUTHERN OHIO	WORK ON UNIT NO. 6 HAS BEEN DELAYED FOR ABOUT A YEAR. CONSTRUCTION IS NOW SCHEDULED TO START JANUARY 1977. UNIT NOW SCHEDULED FOR COMPLETION IN JANUARY 1978
COLUMBIAVILLE NO 6	375 MW - NEW	
COAL	UNIVERSAL OIL PRODUCTS	
LIME SCRUBBING		
STARTUP 2/78		
1.0. NUMBER 29	CONROCK & WILCOX	MODULE A WAS DOWN 3 TIMES LAST MONTH. DUE TO CLEAN DEPOSITS FROM THE VENTURI THROAT AND TWICE DUE TO BOILER RELATED TROUBLES. DEMISTER CONDI- TION DETERIORATED SLIGHTLY AND WILL HAVE TO BE HAND WASHED DURING SCHED- ULED BOILER OUTAGE. OPERATION MOST OF THE MONTH WAS WITH A MEDIUM SULFUR COAL BLEND. AVAILABILITY WAS 85 PERCENT. MODULE B IS STILL DOWN.
107 MW - RETROFIT	COAL 0.6-3.0 PERCENT SULFUR	
CONROCK & WILCOX	LIMESTONE SCRUBBING	
STARTUP 2/72		
1.0. NUMBER 30	DAIRYLAND POWER COOP	EXPERIMENTAL WET LINE FURNACE INJECTION SYSTEM WITH NO TAIL END SCRUBBING IS USED. OUTSTANDING PROBLEMS ARE FURNACE TUBE PLUGGAGE WHICH NECESSI- TATES MANUAL CLEANUP AND SHUT OUT LINE GRINDING AND INJECTION EQUIPMENT. REMOVAL EFFICIENCY WAS FROM 25 TO 40 PERCENT. THE FGD SYSTEM DID NOT RUN MORE THAN TWO DAYS AT A TIME. AVAILABILITY SINCE STARTUP NEVER EXCEEDED 25 PERCENT.
60 MW - RETROFIT	COAL 3.0- 3.5 PERCENT SULFUR	
FOSTER WHEELER	LIME INJECTION	
STARTUP 6/71		

Table 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWDER STATION	CURRENT MONTH
1.0. NUMBER 31		
DETROIT Edison		INSTALLATION OF THE FGD SYSTEM IS ESSENTIALLY COMPLETE. A FAULTY INSTRUMENT PANEL WHICH WAS INCORRECTLY WIRED HAS BEEN RETURNED TO THE MANUFACTURER FOR REPAIR. IT'S REDELIVERY IS NOT EXPECTED BEFORE NOVEMBER 15TH.
ST. CLAY NO. 8		IN THE MEANTIME, DETROIT Edison HAS WATER TESTED SUCH EQUIPMENT AS PUMPS, TANKS AND PIPING. ALSO THE I.O. FANS HAVE BEEN TESTED ON AIR AND THEIR BALANCE CHECKED. STARTUP IS EXPECTED BEFORE THE END OF DECEMBER.
COAL 3.7 PERCENT SULFUR		
PEABODY ELECTRIC		
LIMESTONE SCRUBBER		
STARTUP 11/74		
1.0. NUMBER 32		
MOOREHEAD LIGHT		EQUIPMENT INSTALLED CONSISTS OF FIVE SINGLE STAGE VENTURIS FOR PARTICULATE REMOVAL. ADDITION OF SECOND STAGE VENTURIS OR POSSIBLY JUST USE OF A HIGHLY REACTIVE LIME TO CONTROL SO ₂ EMISSIONS AWAITS RESULTS OF TESTS AT PHILLIPS.
COAL 1.0 - 2.0 PERCENT SULFUR		
CHENICO		
LIME SCRUBBER		
STARTUP 5/75		
1.0. NUMBER 33		
DETROIT Edison		FGD SYSTEM CONSISTS OF 1 TWO STAGE AND 3 ONE STAGE VENTURI SCRUBBERS. AT PRESENT, ONLY FLUE GAS FROM BOILERS 2, 3 AND 4 (ABOUT 40% OF THE STATION CAPACITY) ARE IN SERVICE BECAUSE FLY ASH FROM INEFFICIENT ESP'S OVERLOADS THE CLARIFIERS. OPERATING HOURS FOR MODULES 1 THROUGH 4 BETWEEN 3/17/74 AND 9/30/74 WERE 1756, 762, 815, 1707 RESPECTIVELY. PLANT TRIES TO RUN MODULES 1 AND 4 CONTINUOUSLY.
COAL 1.0 - 2.0 PERCENT SULFUR		
CHENICO		
LIME SCRUBBER		
STARTUP 7/73		
1.0. NUMBER 34		
GENERAL MOTORS		THE FGD SYSTEM HAS BEEN DOWN SINCE OCTOBER 4TH AND WILL REMAIN DOWN TILL NOVEMBER 4TH. THE SHUT DOWN WAS DUE TO TROUBLE IN THE CLARIFIERS RAKE MECHANISM. THE RAKE'S TURBINE GEAR DRIVE WAS DAMAGED BY SOLIDS BUILDUP IN THE GEAR BOX. REPAIRS TO THE CLARIFIER AS WELL AS TO OTHER SIMILAR PROBLEMS TO THE SYSTEM WILL BE MADE DURING THIS SHUTDOWN. THE FGD SYSTEM IS RUNNING AT HALF CAPACITY (SINCE MARCH) DUE TO LOW HEAT DEMANDS.
CHEVROLEL PAPER 1 2 3 & 4		
COAL 2.5 PERCENT SULFUR		
MOOREHEAD LIGHT		
DOUBLE ALKALI		
STARTUP 3/74		
1.0. NUMBER 35		
GENERAL PUBLIC UTILITIES		FIVE PROPOSALS FOR FGD SYSTEMS HAVE BEEN RECEIVED AND A DECISION WILL BE MADE SOON. LIME AND LIMESTONE SCRUBBERS ARE THE FRONT RUNNERS BUT COAL DESULFURIZATION IS ALSO BEING CONSIDERED.
MOOREHEAD LIGHT NO. 3		
COAL 2.0 PERCENT SULFUR		
NOT SELECTED		
STARTUP 4/77		

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

POWER STATION	CURRENT MONTH
1.0. NUMBER 36 ILLINOIS POWER WOOD FIVE 10 4	THE FGD UNIT HAS BEEN DOWN THE PAST 60 DAYS TO PERFORM SEVERAL REPAIRS. THE REPAIR OF THE REFRACTORY BRICKS IN THE HEATER DOME HAS BEEN COMPLETED. REPAIRS TO REFRACTORY LINED DUCTS IS IN PROGRESS. OTHER MAJOR REPAIRS CONSIST OF REINFORCEMENT OF THE STRUCTURAL STEEL SUPPORT TO THE HIGH TEMPERATURE STAINLESS STEEL DUCTWORK AND REPAIR OF THE ACID COOLER DRAFFLES. THE UNIT IS SCHEDULED TO RESTART BY NOVEMBER 22, 1974.
COAL 2.9- 3.2 PERCENT SULFUR NONSULFID FLYING ASH SYSTEMS CATALYTIC OXIDATION STARTUP 10/72	
1.0. NUMBER 37 INDIANAPOLIS POWER AND LIGHT PLEASANT 10 3	INDIANAPOLIS POWER AND LIGHT HAS MET WITH SOME VENDORS IN AN EFFORT TO OBTAIN MORE COMPLETE BIDS. THEY ARE NOW AWAITING REPLIES.
COAL 3.0-3.5 PERCENT SULFUR NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 4/77	
1.0. NUMBER 39 KANSAS CITY POWER & LIGHT HAWTHORNE 10 5	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TO OCTOBER 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, BECAUSE OF THE MANPOWER SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT RESTARTED OCTOBER 1ST AND HAS BEEN RUNNING INTERMITTENTLY DURING THIS MONTH WITH AN AVAILABILITY OF ABOUT 40 PERCENT.
COAL 1.0- 3.0 PERCENT SULFUR CORROSION ENGINEERING LIMESTONE INJECTION ASH SCRUB STARTUP 11/72	
1.0. NUMBER 39 KANSAS CITY POWER & LIGHT HAWTHORNE 10 4	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JUL 8TH TILL OCT 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, DUE TO THE MANPOWER SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT HAS NOT RESTARTED BECAUSE THE COILER IS DOWN TO OVERHAUL THE TURBINE ON THE UNIT'S ELECTRIC GENERATOR.
COAL 2.0- 3.0 PERCENT SULFUR CORROSION ENGINEERING LIMESTONE INJECTION ASH SCRUB STARTUP 8/72	
1.0. NUMBER 40 KANSAS CITY POWER & LIGHT LA CIGAL 10 1	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TILL OCT. 1ST. DURING THIS PERIOD THE BOILER WAS OPERATED BY THE COMPANY'S SUPERVISORY STAFF. THE FGD SYSTEM OPERATED WITH ABOUT 80 PERCENT AVAILABILITY DURING OCTOBER. THERE ARE NO MAJOR OUTSTANDING PROBLEMS. HOWEVER, PREVENTIVE MAINTENANCE PRACTICE REQUIRES THAT ONE MODULE BE TAKEN DOWN ONCE A WEEK FOR 8 HOURS OF CLEANING AND WASHING OF SOLIDS BUILDUP.
COAL 5.2 PERCENT SULFUR BARCOCK & ALLCOX LIMESTONE SCRUBBING STARTUP 6/73	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
1.0. NUMBER 41		
KANSAS POWER & LIGHT		TENTATIVE AGREEMENT HAS BEEN REACHED WITH COMBUSTION ENGINEERING TO DO
JEFFERY CO. 1		PRELIMINARY FGD SYSTEM DESIGN FOR THIS UNIT AS PART OF THEIR CONTINUED
700 MW - NEW		WORK ON THE FGD SYSTEM AT THE LAWRENCE PLANT.
COAL		
COMBUSTION ENGINEERING		
LIVESTONE SCORRING		
STARTUP 6/70		
1.0. NUMBER 42		
KANSAS POWER & LIGHT		TENTATIVE AGREEMENT HAS BEEN REACHED WITH COMBUSTION ENGINEERING TO DO
JEFFERY CO. 2		PRELIMINARY FGD SYSTEM DESIGN FOR THIS UNIT AS PART OF THEIR CONTINUED
700 MW - NEW		WORK ON THE FGD SYSTEM AT THE LAWRENCE PLANT.
COAL		
COMBUSTION ENGINEERING		
LIVESTONE SCORRING		
STARTUP 6/79		
1.0. NUMBER 43		
KANSAS POWER & LIGHT		BOTH FGD MODULES ARE IN OPERATION WITH ESSENTIALLY 100 PERCENT AVAIL-
LAURENCE CO. 1		ABILITY. MAJOR TUNING-UP OF DEVICES IS PERFORMED ABOUT ONCE EVERY TWO
125 MW - RETROFIT		WEEKS. MAJOR CLEANUP OF HEATER COIL IS NOW DUE AFTER MORE THAN 12
COAL 3.0 PERCENT SULFUR		MONTHS OF OPERATION.
COMBUSTION ENGINEERING		
LIVESTONE INJECTION & WET SCRUB		
STARTUP 12/66		
1.0. NUMBER 44		
KANSAS POWER & LIGHT		BOILER 5 HAS BEEN OPERATING ON COAL SINCE OCTOBER 12TH. POOR GAS DISTRI-
LAURENCE CO. 2		BUTION THROUGH EACH ONE OF THE 8 MODULES AS WELL AS AMONG THE MODULES IS
400 MW - NEW		STILL A PROBLEM. COMBUSTION ENGINEERING HAS COMPLETED GAS DISTRIBUTION
COAL 3.0 PERCENT SULFUR		TESTS AND IS CURRENTLY MODIFYING THE GAS FLOW PATTERN THROUGH THE MODULES.
COMBUSTION ENGINEERING		NO RECORD IS KEPT OF THE FGD SYSTEM'S OPERATING HOURS OR SYSTEM AVAIL-
LIVESTONE INJECTION & WET SCRUB		ABILITY.
STARTUP 11/71		
1.0. NUMBER 45		
KUTVICK UTILITIES		THE FGD SYSTEM WILL CONSIST OF ONE MODULE TO HANDLE THE COMBINED FUE-
GREEN RIVER 1 & 2 & 3		GASES FROM THE THREE BOILERS. MAJOR DELAYS WERE EXPERIENCED IN THE
69 MW - RETROFIT		DELIVERY OF STRUCTURAL STEEL. CONSTRUCTION STARTED ABOUT SEPTEMBER 1ST
COAL 3.0 PERCENT SULFUR		AND COMPLETION IS DUE BY MARCH 1975.
AMERICAN AIR FILTER		
LIVE SCORRING		
STARTUP 4/75		

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
100 % - RETROFIT	COAL 3.5- 4.0 PERCENT SULFUR	
NOT SELECTED	LINE SCHEDULING	
STARTUP 6/80		

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/79

COMPANY	POWER STATION	CURRENT MONTH
LOUISVILLE GAS & ELECTRIC MILL CREEK NO. 3	I.O. NUMBER 51	A PERMIT FOR CONSTRUCTION OF THE FGD SYSTEM WAS DISAPPROVED BY THE KENTUCKY PUBLIC SERVICE COMMISSION. HOWEVER, THE ISSUE HAS NOT BEEN FINALLY RESOLVED.
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
	LINE SCRAMBLING STARTUP 6/77	
LOUISVILLE GAS & ELECTRIC MILL CREEK NO. 3	I.O. NUMBER 52	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/79 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
	LINE SCRAMBLING STARTUP 6/77	
LOUISVILLE GAS & ELECTRIC MILL CREEK NO. 3	I.O. NUMBER 53	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/79 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
	LINE SCRAMBLING STARTUP 6/77	
LOUISVILLE GAS & ELECTRIC MILL CREEK NO. 3	I.O. NUMBER 54	CONTRACT WAS AWARDED TO AMERICAN AIR FILTER. SYSTEM WILL OPERATE CLOSED LOOP ON WATER AND SLUDGE WILL BE STABILIZED.
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
	LINE SCRAMBLING STARTUP 6/77	
LOUISVILLE GAS & ELECTRIC MILL CREEK NO. 3	I.O. NUMBER 55	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/79 AS THE STARTUP DATE FOR AN FGD SYSTEM.
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
	LINE SCRAMBLING STARTUP 6/79	

Table 2
STATUS OF FGD SYSTEMS DURING 1974

COMPANY	POWER STATION	CURRENT MONTH
LOUISVILLE GAS & ELECTRIC	56	FGD SYSTEM OPERATING SATISFACTORILY THE PAST 3 WKS ON LIMESTONE UNIT OFF ON ALKALANES. THESE ROUS ARE MADE FOR AND FUNDED BY COMBUSTION ENGINEERING. AVAILABILITY WAS ESSENTIALLY 100 PERCENT WITH NO PROBLEMS EXPERIENCED IN ANY PART OF THE SYSTEM. THE LIMESTONE SLODGE IS DEWATERED IN THE EXISTING ROTARY VACUUM FILTERS WITH NO BLINDING OF FILTER CLOTH.
COAL 3.5-4.0 PERCENT SULFUR		
COMBUSTION ENGINEERING		
LIME SCHEDULING		
STARTUP 4/74		
LOUISVILLE GAS & ELECTRIC	57	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. ANTICIPATE COMPLETING CONSTRUCTION ON SCHEDULE.
MONTANA POWER		
COLSTRIP NO. 1		
300 MW - NEW		
COAL 0.6 PERCENT SULFUR		
COMBUSTION EQUIP. ASSOCIATES		
LIME SCHEDULING		
STARTUP 5/75		
LOUISVILLE GAS & ELECTRIC	58	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. ANTICIPATE COMPLETING CONSTRUCTION ON SCHEDULE.
MONTANA POWER		
COLSTRIP NO. 2		
300 MW - NEW		
COAL 0.6 PERCENT SULFUR		
COMBUSTION EQUIP. ASSOCIATES		
LIME SCHEDULING		
STARTUP 5/75		
LOUISVILLE GAS & ELECTRIC	59	UTILITY IS WAITING FOR APPROVAL OF A CERTIFICATE TO CONSTRUCT BOILERS NO. 3 AND NO. 4 BEFORE PROCEEDING WITH PLANS TO INSTALL FGD SYSTEMS.
MONTANA POWER		
COLSTRIP NO. 3		
700 MW - NEW		
COAL 0.7 PERCENT SULFUR		
NOT SELECTED		
NOT SELECTED		
STARTUP 7/76		
LOUISVILLE GAS & ELECTRIC	60	UTILITY IS WAITING FOR APPROVAL OF A CERTIFICATE TO CONSTRUCT BOILERS NO. 3 AND NO. 4 BEFORE PROCEEDING WITH PLANS TO INSTALL FGD SYSTEMS.
MONTANA POWER		
COLSTRIP NO. 4		
700 MW - NEW		
COAL 0.7 PERCENT SULFUR		
NOT SELECTED		
NOT SELECTED		
STARTUP 0/79		

TABLE 2
STATUS OF FOD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
NEVADA POWER REID GARDNER NO 1	125 MW - RETROFIT	FGO SYSTEM PERFORMANCE IS SATISFACTORILY WITH NO OUTSTANDING MECHANICAL OR PROCESS PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADEQUATE SUPPLY OF TRONA (SODIUM CARBONATE GRI) WHICH HAS RESULTED IN FREQUENT INTERRUPTIONS IN THE OPERATION OF THE FGO SYSTEM. NEVADA POWER IS SUEING THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SODIUM CARBONATE BRINE PURCHASED FROM AN ALTERNATE SUPPLIER.
NEVADA POWER REID GARDNER NO 2	125 MW - RETROFIT	FGO SYSTEM PERFORMANCE IS SATISFACTORILY WITH NO OUTSTANDING MECHANICAL OR PROCESS PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADEQUATE SUPPLY OF TRONA (SODIUM CARBONATE GRI) WHICH HAS RESULTED IN FREQUENT INTERRUPTIONS IN THE OPERATION OF THE FGO SYSTEM. NEVADA POWER IS SUEING THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SODIUM CARBONATE BRINE PURCHASED FROM AN ALTERNATE SUPPLIER.
NEVADA POWER REID GARDNER NO 3	125 MW - RETROFIT	FGO SYSTEM IS UNDER CONSTRUCTION.
NEVADA POWER REID GARDNER NO 4	125 MW - RETROFIT	NEVADA POWER COMPANY HAS SIGNED A LETTER OF INTENT WITH CORUSION EQUIPMENT ASSOCIATES FOR THE CONSTRUCTION OF AN FGO SYSTEM ON REID GARDNER NO. 4. THIS UNIT IS SCHEDULED FOR COMPLETION IN 1978.
NEVADA POWER REID GARDNER NO 5	125 MW - RETROFIT	UNIVERSAL OIL PRODUCTS IS CONDUCTING PRELIMINARY DESIGN WORK ON A DRY FGO UNIT WITH SULFUR RECOVERY. DECISION TO BE MADE AROUND NOV 1 WHETHER TO SIGN AN ENGINEERING CONTRACT.

TABLE 2
STATUS OF FGD SYSTEMS DURING 1974

COMPANY	POWER STATION	CURRENT MONTH
NORTHERN INDIANA PUB SERVICE	100 TONNER 69	CONSIDERING A LINE ON LIMESTONE SCRUBBING UNIT. ALSO WAITING FOR PERFORMANCE OF WELLMAN LONG UNIT
EMERY CO.	190 VA - RETROFIT	UNDER CONSTRUCTION AT THEIR D.M. MITCHELL NO.11 UNIT
COAL 3 PERCENT SULFUR		
NOT SELECTED		
NOT SELECTED		
STARTUP 6/3		
NORTHERN INDIANA PUB SERVICE	100 TONNER 67	CONSIDERING A LINE ON LIMESTONE SCRUBBING UNIT. ALSO WAITING FOR PERFORMANCE OF WELLMAN LONG UNIT
DAILY NO. 4	400 VA - RETROFIT	UNDER CONSTRUCTION AT THEIR D.M. MITCHELL NO.11 UNIT
COAL 3 PERCENT SULFUR		
NOT SELECTED		
NOT SELECTED		
STARTUP 6/3		
NORTHERN INDIANA PUB SERVICE	100 TONNER 68	SULFUR DIOXIDE WILL BE CONVERTED TO SULFURIC ACID IN ADJACENT PLANT USING THE ALLIED CHEMICAL PROCESS. THIS INSTALLATION IS PARTIALLY FUNDED BY AN EPA DEMONSTRATION GRANT.
MITCHELL NO. 11	115 VA - RETROFIT	
COAL 3.2- 3.5 PERCENT SULFUR		
DAVE FOREMAN		
WELLMAN LONG		
STARTUP 6/75		
NORTHERN STATES POWER	100 TONNER 69	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. PURCHASE ORDERS FOR SCRUBBERS WERE PLACED IN EARLY 1974. RECENT SCRUBBER DESIGN CHANGES HAVE CAUSED EQUIPMENT DELIVERY DELAYS BUT THE OVERALL SCHEDULE HAS NOT BEEN UPSET. PROBLEMS HAVE ARISEN IN OBTAINING SOME ELECTRICAL CONTROLS.
SHELBOURNE NO. 1	600 VA - 12"	
COAL 1.0 PERCENT SULFUR		
CONSTRUCTION ENGINEERING		
LIVESTONE SCRUBBING		
STARTUP 5/76		
NORTHERN STATES POWER	100 TONNER 70	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. PURCHASE ORDERS FOR SCRUBBERS WERE PLACED IN EARLY 1974. RECENT SCRUBBER DESIGN CHANGES HAVE CAUSED EQUIPMENT DELIVERY DELAYS BUT THE OVERALL SCHEDULE HAS NOT BEEN UPSET. PROBLEMS HAVE ARISEN IN OBTAINING SOME ELECTRICAL CONTROLS.
SHELBOURNE NO. 2	600 VA - 12"	
COAL 3.0 PERCENT SULFUR		
CONSTRUCTION ENGINEERING		
LIVESTONE SCRUBBING		
STARTUP 5/77		

TABLE 2

STATUS OF FGD SYSTEMS DURING 10/74

COMPANY

POWER STATION

CURRENT MONTH

I.D. NUMBER 71

PENNSYLVANIA POWER CO.

BRUCE MANFIELD NO. 1

FGD NO. - NEW

COAL 4.3 PERCENT SULFUR

CHEMICO

LIME SCRUBBING

STARTUP 8/75

I.D. NUMBER 72

PENNSYLVANIA POWER CO.

BRUCE MANFIELD NO. 2

FGD NO. - NEW

COAL 4.3 PERCENT SULFUR

CHEMICO

LIME SCRUBBING

STARTUP 8/75

I.D. NUMBER 73

PENNSYLVANIA POWER CO.

BRUCE MANFIELD NO. 3

FGD NO. - NEW

COAL 4.3 PERCENT SULFUR

NOT SELECTED

LIME SCRUBBING

STARTUP 8/74

I.D. NUMBER 74

PHILADELPHIA ELECTRIC

EDDYSTONE NO. 1

120 NO. - RETROFIT

COAL 2.5 PERCENT SULFUR

UNITED ENGINEERS

MAGNESIUM OXIDE SCRUBBING

STARTUP 12/74

I.D. NUMBER 75

POTOMAC ELECTRIC & POWER

WICKESBORO NO. 3

100 NO. - RETROFIT

COAL 2.0 PERCENT SULFUR

CHEMICO

MAGNESIUM OXIDE SCRUBBING

STARTUP 9/75

IN PLANNING STAGE

THE FGD SYSTEM CONSISTS OF 3 PARALLEL VENTURI SCRUBBERS (ONLY ONE IS FOLLOWED BY A 2-STAGE VENTURI SCRUBBER FOR SO₂ REMOVAL). TIE IN OF MODULES TO THE BOILER HAS STARTED AND IS DUE FOR COMPLETION BY DECEMBER 1ST. THE SYSTEM WILL BE TESTED WITH WATER FOR PARTICULATE REMOVAL AND DEBUGGING DURING DECEMBER AND JANUARY. SO₂ SCRUBBING WILL NOT BEGIN BEFORE FEBRUARY 1ST.

THE FGD SYSTEM CONSISTS OF ONE VENTURI STAGE FOR FLY ASH SCRUBBING WITH WATER FOLLOWED BY A SECOND VENTURI STAGE FOR SO₂ SCRUBBING WITH WASH SLURRY. THE SYSTEM WAS OPERATED FROM OCTOBER 3RD TO THE 14TH. PRESENTLY THE SYSTEM IS DOWN TO REPAIR LEAKS IN THE FIRST VENTURI STAGE AS WELL AS PROBLEMS WITH THE MGS03 BUCKET ELEVATOR. AVAILABILITY SINCE AUGUST 20TH AVERAGED 34 PERCENT.

COMPANY		Table 2 STATUS OF FGD SYSTEMS DURING 10/74	
POWER STATION	CURRENT MONTH		
100. NUMBER 76			
PUBLIC SERVICE OF INDIANA			UNIT NO. 2 SHOULD BE COMPLETED BY SUMMER 1975. UTILITY IS EXPERIENCING SOME DIFFICULTY IN OBTAINING CONSTRUCTION MATERIALS AND CHEMICALS.
GLISON NO. 2			TOTAL UNIT CAPACITY IS 650 Mw.
650 Mw - RETROFIT			
COAL 0.8 PERCENT SULFUR			
NOT SELECTED			
LIME/LIMESTONE SCRUBBING			
STARTUP 5/75			
100. NUMBER 77			
PUBLIC SERVICE OF NEW MEXICO			FGD PROJECT IS PROCEEDING ACCORDING TO PLAN WITH NO MAJOR PROBLEMS.
SAN JUAN NO. 2			EQUIPMENT IS PRESENTLY BEING PROCURED. CONSTRUCTION TO BEGIN IN AUGUST 1975
375 Mw - NEW			
COAL 0.8 PERCENT SULFUR			
DAVY POWERGAS			
WELLMAN CORP.			
STARTUP 1/77			
100. NUMBER 78			
PUBLIC SERVICE OF NEW MEXICO			FGD PROJECT IS PROCEEDING ACCORDING TO PLAN WITH NO MAJOR PROBLEMS.
SAN JUAN NO. 2			EQUIPMENT IS PRESENTLY BEING PROCURED. CONSTRUCTION TO BEGIN IN AUGUST 1975
390 Mw - RETROFIT			
COAL 0.8 PERCENT SULFUR			
DAVY POWERGAS			
WELLMAN CORP.			
STARTUP 1/77			
100. NUMBER 79			
PUBLIC SERVICE OF NEW MEXICO			PLANS FOR THE BOILER ARE IN THE ENGINEERING STAGE. WORK ON FGD SYSTEM IS IN THE EARLY PLANNING STAGE.
SAN JUAN NO. 2			
500 Mw - NEW			
COAL 0.8 PERCENT SULFUR			
NOT SELECTED			
NOT SELECTED			
STARTUP 5/76			
100. NUMBER 80			
PUBLIC SERVICE OF NEW MEXICO			PLANS FOR THE BOILER ARE IN THE ENGINEERING STAGE. WORK ON FGD SYSTEM IS IN THE EARLY PLANNING STAGE.
SAN JUAN NO. 2			
500 Mw - RETROFIT			
COAL 0.8 PERCENT SULFUR			
NOT SELECTED			
NOT SELECTED			
STARTUP 5/80			

PRINTED IN U.S.A.

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
S. CAROLINA PUB SERV AUTHORITY GEORGETOWN NO. 2 140 MW - NEW	I.D. NUMBER 81	VENDOR NOT SELECTED. BIDS ARE DUE IN OCTOBER 74. THE STATION CAPACITY IS 200 MW. ALTERNATIVE BIDS HAVE BEEN REQUESTED FOR THE TOTAL 200 MW CAPACITY AND A 1.9% SULFUR COAL.
	NOT SELECTED	
	LIME/LIMESTONE SCRUBBING	
	STARTUP 5/77	
SALT RIVER PROJECT NAVAJO NO. 1 750 MW - NEW	I.D. NUMBER 82	AWAITING TEST RESULTS ON HORIZONTAL AND VERTICAL SCRUBBING SYSTEMS AT MONAHE. CONTRACT WILL BE AWARDED 6/1/75.
	NOT SELECTED	
	LIME/LIMESTONE SCRUBBING	
	STARTUP 3/76	
SALT RIVER PROJECT NAVAJO NO. 2 750 MW - NEW	I.D. NUMBER 83	AWAITING TEST RESULTS ON HORIZONTAL AND VERTICAL SCRUBBING SYSTEMS AT MONAHE. CONTRACT WILL BE AWARDED 6/1/75.
	NOT SELECTED	
	LIME/LIMESTONE SCRUBBING	
	STARTUP 10/75	
SALT RIVER PROJECT NAVAJO NO. 3 750 MW - NEW	I.D. NUMBER 84	AWAITING TEST RESULTS ON HORIZONTAL AND VERTICAL SCRUBBING SYSTEMS AT MONAHE. CONTRACT WILL BE AWARDED 6/1/75.
	NOT SELECTED	
	LIME/LIMESTONE SCRUBBING	
	STARTUP 3/77	
SOUTHERN CALIFORNIA Edison MONAHE NO. 1 640 MW - REINFORCED	I.D. NUMBER 85	THE TOTAL CAPACITY OF MONAHE IS 600 MW. EXISTING EXPERIMENTAL FGD MODULE WILL TEST ONLY 100 MW. FINAL SELECTION OF THE PROCESS IS PENDING THE OUTCOME OF TEST RUNS NOW IN PROGRESS AT THE MONAHE STATION.
	NOT SELECTED	
	LIME SCRUBBING	
	STARTUP 6/77	

PROJECT NO. 5A

FORM 103

21
22
23
24
25
26
27
28
29
30
31
32
33
34
35
36
37
38
39
40
41
42
43
44
45
46
47
48
49
50
51
52
53
54
55
56
57
58
59
60
61
62
63
64
65
66
67
68
69
70
71
72
73
74
75
76
77
78
79
80
81
82
83
84
85
86
87
88
89
90
91
92
93
94
95
96
97
98
99
100

Table 2
STATUS OF FGD SYSTEMS DURING 1974

COMPANY	POWER STATION	CURRENT MONTH
SOUTHERN CALIFORNIA Edison	100 MW - NO	THE ASPHALT WHICH WERE PLACED FOR THE FIVE DAMAGED EQUIPMENT ON THE VERTICAL MODULE ARE ESSENTIALLY COMPLETE. THE SYSTEM HAS BEEN CHECKED OUT USING AMBIENT AIR AND WATER. THE FINAL PHASE OF STARTUP USING FLUE GAS AND LIMESTONE SLURRY IS NOW IN PROGRESS.
	150 MW - RETROFIT	
	COAL 0.5-0.6 PERCENT SULFUR	
	UNIVERSAL OIL PRODUCTS Limestone Scrubbing STARTUP 12/74	
SOUTHERN CALIFORNIA Edison	100 MW - NO	THE TEST RUNS UNDER THE NAVAJO PLANT OPERATING CONDITIONS HAVE BEEN COMPLETED. THIS EXPERIMENTAL HORIZONTAL MODULE HAS BEEN TEST RUN ON LIMESTONE FOR ABOUT TWO WEEKS IN OCTOBER. THE UNIT IS NOW BACK ON LINE. THE AVERAGE AVAILABILITY OF THIS FGD SYSTEM SINCE ITS INITIAL START UP IN JANUARY 1974 IS ABOUT 80 PERCENT.
	150 MW - RETROFIT	
	COAL 0.5-0.6 PERCENT SULFUR	
	UNIVERSAL OIL PRODUCTS Limestone Scrubbing STARTUP 11/75	
SOUTHERN CALIFORNIA Edison	100 MW - NO	THE TOTAL CAPACITY OF BOILER IS 800 MW. EXISTING EXPERIMENTAL FGD MODULE WILL TREAT ONLY 160 MW. FINAL SELECTION OF THE PROCESS IS PENDING THE OUTCOME OF TEST RUNS NOW IN PROGRESS AT THE NAVAJO STATION.
	150 MW - RETROFIT	
	COAL 0.5-0.6 PERCENT SULFUR	
	UNIVERSAL OIL PRODUCTS Limestone Scrubbing STARTUP 5/77	
SOUTHERN MISSISSIPPI PWR COOP	100 MW - NO	FGD SYSTEM IS IN EARLY PLANNING STAGE. UTILITY COMPANY'S CONSULTANT, BURNS & MCDONNELL, IS PRESENTLY PREPARING SPECIFICATIONS FOR BIDS DOCUMENTS.
	150 MW - RETROFIT	
	COAL 0.5-0.6 PERCENT SULFUR	
	UNIVERSAL OIL PRODUCTS Limestone Scrubbing STARTUP 6/79	
SOUTHERN MISSISSIPPI PWR COOP	100 MW - NO	FGD SYSTEM IS IN EARLY PLANNING STAGE. UTILITY COMPANY'S CONSULTANT, BURNS & MCDONNELL, IS PRESENTLY PREPARING SPECIFICATIONS FOR BIDS DOCUMENTS.
	150 MW - RETROFIT	
	COAL 0.5-0.6 PERCENT SULFUR	
	UNIVERSAL OIL PRODUCTS Limestone Scrubbing STARTUP 6/79	

TABLE 2
STATUS OF FGD SYSTEMS DURING 1974

COMPANY	POWER STATION	CURRENT MONTH
SOUTHERN POWER CO.	1.0. NUMBER 91 SOUTHERN POWER CO. PUBLIC SERVICE HARRINGTON RD 1 343 P.A. - NEW COAL 0.5 PERCENT SULFUR COMBUSTION ENGINEERING LIME/LIMESTONE SCRUBBING STARTUP 8/75	PLANT IS UNDER CONSTRUCTION BUT EXPERIENCING DELAYS IN DELIVERY OF STRUCTURAL STEEL AND ALLOY PIPES AND VALVES. EXPECT TO HAVE STRUCTURAL STEEL FOR SCRUBBERS ERECTED BY 3RD QUARTER 1975.
AMERICAN ELECTRIC CO.	1.0. NUMBER 92 SPRINGFIELD UTILITY BOARD SOUTHWEST RD 1 200 P.A. - NEW COAL UNIVERSAL OIL PRODUCTS LIMESTONE SCRUBBING STARTUP 6/76	THE BOILER IS UNDER CONSTRUCTION. INSTALLATION OF FGD SYSTEM WILL START IN NOVEMBER 74. PRESENTLY, THE FGD PROJECT IS IN ENGINEERING AND EQUIPMENT PROCUREMENT STAGE.
AMERICAN ELECTRIC CO.	1.0. NUMBER 93 TENNESSEE VALLEY AUTHORITY SHAWNEE RD. 10 30 P.A. - RETROFIT COAL TENNESSEE VALLEY AUTHORITY LIME/LIMESTONE SCRUBBING STARTUP 4/72	THIS IS AN EXPERIMENTAL SYSTEM FUNDED BY USEPA. THERE ARE THREE TYPES OF SCRUBBERS (10 P.A. EQUIVALENT EACH) BEING EVALUATED UNDER A VARIETY OF OPERATIONAL MODES.
AMERICAN ELECTRIC CO.	1.0. NUMBER 94 TENNESSEE VALLEY AUTHORITY ALCOA GREEN RD 4 950 P.A. - RETROFIT COAL 0.7 PERCENT SULFUR TENNESSEE VALLEY AUTHORITY LIMESTONE SCRUBBING STARTUP 1/77	ALL MAJOR PROCESS EQUIPMENT HAS BEEN PURCHASED. IN-LINE REHEATER SCRAPPED AND DIRECT HEAT SYSTEM IS IN DESIGN STAGE.
AMERICAN ELECTRIC CO.	1.0. NUMBER 95 TEXAS GASLIGHTS MARTIN LAKE RD. 1 793 P.A. - NEW COAL 0.4 PERCENT SULFUR RESEARCH COTTELL LIMESTONE SCRUBBING STARTUP 9/76	NEW SYSTEM, ERECTING BOILER. CONTRACT FOR FGD SYSTEM AWARDED.

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

CURRENT MONTH

POWER STATION

100 NUMBER 96

TEXAS UTILITIES

MARTIN LANE 10.2

793 PW - NEW

COAL 0.44 PERCENT SULFUR

RESEARCH COTINELL

LIMESTONE SCRUBBING

STARTUP 6/77

NEW SYSTEM, ERECTING BOILER. CONTRACT FOR FGD SYSTEM AWARDED.

100 NUMBER 97

TEXAS UTILITIES

MARTIN LANE 10.3

793 PW - NEW

COAL 0.44 PERCENT SULFUR

RESEARCH COTINELL

LIMESTONE SCRUBBING

STARTUP 12/78

NEW SYSTEM, ERECTING BOILER. COMPANY IS OUT FOR BIDS ON FGD SYSTEM.

100 NUMBER 98

TEXAS UTILITIES

MARTIN LANE 10.4

793 PW - NEW

COAL 0.44 PERCENT SULFUR

RESEARCH COTINELL

LIMESTONE SCRUBBING

STARTUP 12/78

NEW SYSTEM, ERECTING BOILER. COMPANY IS OUT FOR BIDS ON FGD SYSTEM.

100 NUMBER 99

TEXAS UTILITIES

MONTICELLO 10.3

800 PW - NEW

COAL 0.44-1.0 PERCENT SULFUR

NOT SELECTED

LIMESTONE SCRUBBING

STARTUP 12/78

INVITATION FOR BIDS ON FGD SYSTEM HAVE BEEN SENT OUT AND SOME PROPOSALS HAVE BEEN RECEIVED. THE COMPANY IS PRESENTLY STUDYING AND EVALUATING THESE PROPOSALS.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	17
UTILITY NAME	BOSTON Edison
UNIT NAME	MYSTIC NO 6
UNIT LOCATION	EVERETT MASSACHUSETTS
UNIT RATING	150 MW
FUEL CHARACTERISTICS	OIL 2.5 PERCENT SULFUR
FGD VENDOR	CHERICO
PROCESS	ABSORPTION OXIDE SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	4/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	50 PERCENT
SO ₂	90 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	
UNIT COST	\$25/KW CAPITAL \$0.9 MILLS/KWH OP

OPERATIONAL EXPERIENCE FGD SYSTEM IS USED WITH NO IMMEDIATE PLANS FOR RESTARTING BECAUSE EPA FUNDING HAS EXPIRED AND THE CALCINING PLANT IS NOW BEING USED BY ANOTHER COMPANY. THE UTILITY IS EVALUATING THE DATA COLLECTED DURING THE DEMONSTRATION PERIOD TO DETERMINE THE COURSE OF FUTURE ACTION.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	22
UTILITY NAME	CITY OF KEY WEST
UNIT NAME	KEY WEST POWER PLANT
UNIT LOCATION	KEY WEST FLORIDA
UNIT RATING	37 MW
FUEL CHARACTERISTICS	OIL 2.4 PERCENT SULFUR
FGD VENDOR	ZURN AIR SYSTEMS
PROCESS	LIMESTONE SCRUBBING
NEW OR RETROFIT	NEW
START UP DATE	10/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	90 PERCENT
SO2	85 PERCENT-ESTIMATED
WATER MAKE UP	MAKE UP RATE ESTIMATED 100 GPM
SLUDGE DISPOSAL	UNLINED POND
UNIT COST	

OPERATIONAL EXPERIENCE
PLANT HAS BEEN SHUT DOWN SINCE SEPT 19TH BECAUSE OF A BOILER I D FAN MO-
TOR FAILURE. THEY HOPE TO BE OPERATIONAL THE WEEK OF OCTOBER 21ST. MECH-
ANICAL FAILURES IN THE FGD SYSTEM WERE FIXED. FGD OPERATED ABOUT A WEEK
AND A HALF BEFORE THE PRESENT BOILER SHUTDOWN. DURING SHUTDOWN THE EX-
HAUST JOINTS ON THE DUCTWORK TO THE SCRUBBER WERE REPLACED FOR THE 2ND
TIME. SOME PLUGGING REMOVED AT SHUTDOWN.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	29
UTILITY NAME	COMMONWEALTH EDISON
UNIT NAME	WILL COUNTY NO 1
UNIT LOCATION	ROSEMONT ILLINOIS
UNIT RATING	157 MW
FUEL CHARACTERISTICS	COAL 0.6-3.0 PERCENT SULFUR
FGD VENDOR	BABCOCK & WILCOX
PROCESS	LIMESTONE SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	2/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	96 PERCENT
SO ₂	75-80 PERCENT-AVERAGE
WATER MAKE UP	106 GAL/LB MOLE SO ₂ REMOVED
SLUDGE DISPOSAL	STABILIZED SLUDGE DISPOSED IN CLAY LINED POND
UNIT COST	\$96/KW CAPITAL+\$4.5 MILS/KWH OP

OPERATIONAL EXPERIENCE
MODULE A WAS DOWN 3 TIMES LAST MONTH DUE TO CLEAN DEPOSITS FROM THE VENTURI THROAT AND TWICE DUE TO BOILER RELATED TROUBLES. DENISTER CONTINUED DETERIORATED SLIGHTLY AND WILL HAVE TO BE HAND WASHED DURING SCHEDULED BOILER OUTAGE. OPERATION MOST OF THE MONTH WAS WITH A MEDIUM SULFUR COAL BLEND. AVAILABILITY WAS 65 PERCENT. MODULE B IS STILL DOWN.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	30
UTILITY NAME	DAIRYLAND POWER COOP
UNIT NAME	ALMA STATION
UNIT LOCATION	ALMA WISCONSIN
UNIT RATING	60 Mw
FUEL CHARACTERISTICS	COAL 3.0- 3.5 PERCENT SULFUR
FGD VENDOR	FOSTER WHEELER
PROCESS	LIME INJECTION
NEW OR RETROFIT	RETROFIT
START UP DATE	6/71
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	25 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	
UNIT COST	
OPERATIONAL EXPERIENCE	EXPERIMENTAL WET LIME FURNACE INJECTION SYSTEM WITH NO TAIL END SCRUBBING IS USED. OUTSTANDING PROBLEMS ARE FURNACE TUBE PLUGGING WHICH NECESSITATED MANUAL CLEANUP AND WORN OUT LIME GRINDING AND INJECTION EQUIPMENT. REMOVAL EFFICIENCY WAS FROM 25 TO 40 PERCENT. THE FGD SYSTEM DID NOT RUN MORE THAN TWO DAYS AT A TIME. AVAILABILITY SINCE STARTUP NEVER EXCEEDED 25 PERCENT.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO. 33
UTILITY NAME EUGENE LIGHT
UNIT NAME PHILLIPS
UNIT LOCATION SOUTH HEIGHT PENNSYLVANIA
UNIT RATING 410 MW

FUEL CHARACTERISTICS COAL 1.0- 2.0 PERCENT SULFUR
FGD VENDOR CHEMICO

PROCESS LINE SMOGGING

NEW OR RETROFIT RETROFIT

START UP DATE 7/73

FGD STATUS OPERATIONAL

EFFICIENCY,
PARTICULATES

SO2

WATER MAKE UP

SLUDGE DISPOSAL STABILIZED SLUDGE DISPOSED IN UNLINED POND

UNIT COST \$70-75/KW CAP. \$2.1 MILS/KWH OP

OPERATIONAL
EXPERIENCE FGD SYSTEM CONSISTS OF 1 TWO STAGE AND 3 ONE STAGE VERMICULI SCRUBBERS. AT PRESENT, ONLY FLUE GAS FROM BOILERS 2,3 AND 4 (ABOUT 40% OF THE STATION CAPACITY) ARE IN SERVICE BECAUSE FLY ASH FROM INEFFICIENT ESP'S OVERLOADS THE CLANFILLERS. OPERATING HOURS FOR MODULES 1 THROUGH 4 BETWEEN 3/17/74 AND 8/30/74 WERE 1756, 762, 815, 1707 RESPECTIVELY. PLANT TRIES TO RUN MODULES 1 AND 4 CONTINUOUSLY.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/7/74

IDENTIFICATION NO. 34
UTILITY NAME GENERAL MOTORS
UNIT NAME CHEVROLET FARM 1 2 3 & 4
UNIT LOCATION PARMA OHIO
UNIT RATING 32 MW

FUEL CHARACTERISTICS COAL 2.5 PERCENT SULFUR

FGD VERSION KOCH

PROCESS DOUBLE ALKALI

NEW OR RETROFIT RETROFIT

START UP DATE 3/74

FGD STATUS OPERATIONAL

EFFICIENCY,
PARTICULATES

SO2

WATER MAKE UP

SLUDGE DISPOSAL

UNIT COST

OPERATIONAL
EXPERIENCE

THE FGD SYSTEM HAS BEEN DOWN SINCE OCTOBER 4TH AND WILL REMAIN DOWN TILL NOVEMBER 4TH. THE SHUT DOWN WAS DUE TO TROUBLE IN THE CLARIFIERS HAVE MECHANISM. THE NAME'S TURNING GEAR DRIVE WAS DAMAGED BY SOLIDS BUILDUP IN THE GEAR BOX. REPAIRS TO THE CLARIFIER AS WELL AS TO OTHER MINOR PROBLEMS TO THE SYSTEM WILL BE MADE DURING THIS SHUT DOWN. THE FGD SYSTEM IS RUNNING AT HALF CAPACITY (SINCE MARCH) DUE TO LOW HEAT DEMANDS.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	36
UTILITY NAME	ILLINOIS POWER
UNIT NAME	WOOD RIVER NO 4
UNIT LOCATION	EAST ALTON, ILLINOIS
UNIT RATING	110 MW
FUEL CHARACTERISTICS	COAL 2.9- 3.2 PERCENT SULFUR
FGD VENDOR	MONSIEUR ENVIRONNEMENTAL SYSTEMS
PROCESS	CATALYTIC OXIDATION
NEW OR RETROFIT	RETROFIT
START UP DATE	10/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	85 PERCENT
WATER MAKE UP	
SLODGE DISPOSAL	
UNIT COST	\$73/KW CAPITAL + \$4 MILLS/KWH OP
OPERATIONAL EXPERIENCE	THE FGD UNIT HAS BEEN DOWN THE PAST 60 DAYS TO PERFORM SEVERAL REPAIRS. THE REPAIR OF THE REFRACTORY BRICKS IN THE HEATER BURNER DOME HAS BEEN COMPLETED. REPAIRS TO REFRACTORY LINED DUCTS IS IN PROGRESS. OTHER MAJOR REPAIR CONSIST OF REINFORCEMENT OF THE STRUCTURAL STEEL SUPPORT TO THE HIGH TEMPERATURE STAINLESS STEEL DUCTWORK AND REPAIR OF THE ACID COOLER BAFFLES. THE UNIT IS SCHEDULED TO RESTART BY NOVEMBER 22, 1974.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	36
UTILITY NAME	KANSAS CITY POWER & LIGHT
UNIT NAME	HAATHORN NO 3
UNIT LOCATION	KANSAS CITY MISSOURI
UNIT RATING	140 MW
FUEL CHARACTERISTICS	COAL 0.8- 3.0 PERCENT SULFUR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIMESTONE INJECTION WET SCRUB
NEW OR RETROFIT	RETROFIT
START UP DATE	11/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	99 PERCENT
SO2	70 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$19/KW CAPITAL+2.5 MILS/ANW OP
OPERATIONAL EXPERIENCE	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TO OCTOBER 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, BECAUSE OF THE MANPOWER SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT RESTARTED OCTOBER 1ST AND HAS BEEN RUNNING INTERMITTENTLY DURING THIS MONTH WITH AN AVAILABILITY OF ABOUT 40 PERCENT.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 1974

IDENTIFICATION NO.	39
UTILITY NAME	KANSAS CITY POWER & LIGHT
UNIT NAME	MAIN TOWER NO. 4
UNIT LOCATION	KANSAS CITY, MISSOURI
UNIT RATING	100 MW
FUEL CHARACTERISTICS	COAL 0.8-3.0 PERCENT SULFUR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIMESTONE INJECTION WET SCRUB
NEW OR RETROFIT	RETROFIT
START-UP DATE	8/72
FGD STATUS	OPERATIONAL
EFFICIENCY PARTICULATES	99 PERCENT
SO ₂	70 PERCENT
WATER MAKE-UP	
SLUDGE DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$19/KW CAPITAL \$2.2 MILS/KWH OP
OPERATIONAL EXPERIENCE	ACPL HAD A LABOUR STRIKE AT THE PLANT WHICH LASTED FROM JUL 8TH TILL OCT 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, DUE TO THE KANSAS CITY SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT HAS NOT RESTARTED BECAUSE THE BOILER IS DOWN TO OVERHAUL THE TURBINE ON THE UNIT'S ELECTRIC GENERATOR.

TABLE 2
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	40
UTILITY NAME	KANSAS CITY POWER & LIGHT
UNIT NAME	LA CYCNE NO 3
UNIT LOCATION	LA CYCNE KANSAS
UNIT RATING	820 MW
FUEL CHARACTERISTICS	COAL 5.2 PERCENT SULFUR
FGD METHOD	BAUDOUIN & WILCOX
PROCESS	LIVESTONE SCRUBBING
NEW OR RETROFIT	NEW
START UP DATE	6/73
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	98 PERCENT
SO2	72 PERCENT
WATER MADE UP	
SLOUGE DISPOSAL	UNSTABILIZED SLOUGE DISPOSED IN UNLINED POND
UNIT COST	\$4.37/KW CAPITAL

OPERATIONAL EXPERIENCE
ACPL HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TILL OCT. 1ST. DURING THIS PERIOD THE BOILER WAS OPERATED BY THE COMPANY'S SUPERVISORY STAFF. THE FGD SYSTEM OPERATED WITH ABOUT 60 PERCENT AVAILABILITY DURING OCCASION. THERE ARE NO MAJOR OUTSTANDING PROBLEMS. HOWEVER, PREVENTIVE MAINTENANCE PRACTICE REQUIRES THAT ONE SHOULD BE TAKEN DOWN ONCE A WEEK FOR 8 HOURS OF CLEANING AND WASHING OF SOLIDS BUILDUP.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	43
UTILITY NAME	KANSAS POWER & LIGHT
UNIT NAME	LAWRENCE NO. 4
UNIT LOCATION	LAWRENCE, KANSAS
UNIT RATING	125 MW
FUEL CHARACTERISTICS	COAL 3.5 PERCENT SULFUR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIME/STONE INJECTION WET SCRUB
NEW OR RETROFIT	RETROFIT
START-UP DATE	12/68
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	99 PERCENT
SO ₂	85 PERCENT
WATER MAKE UP	
SLOUDGE DISPOSAL	UNSTABILIZED SLOUDGE DISPOSED IN UNLINED POND
UNIT COST	
OPERATIONAL EXPERIENCE	NO. 4 FGD MODULES ARE IN OPERATION WITH ESSENTIALLY 100 PERCENT AVAILABILITY. ANNUAL WASH-UP OF DEMISTERS IS PERFORMED ABOUT ONCE EVERY TWO WEEKS. ANNUAL CLEANUP OF REHEATER COIL IS NOW DUE AFTER MORE THAN 12 MONTHS OF OPERATION.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	49
UTILITY NAME	KANSAS POWER & LIGHT
UNIT NAME	LAWRENCE NO 5
UNIT LOCATION	LAWRENCE KANSAS
UNIT RATING	400 MW
FUEL CHARACTERISTICS	COAL 3.5 PERCENT SULFUR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIVESTONE INJECTION WET SCRUB
NEW OR RETROFIT	NEW
START UP DATE	11/71
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	99 PERCENT
SO2	65 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	
OPERATIONAL EXPERIENCE	ROLLER 5 HAS BEEN OPERATING ON COAL SINCE OCTOBER 12TH. POOR GAS DISTRIBUTION THROUGH EACH ONE OF THE 8 MODULES AS WELL AS AMONG THE MODULES IS STILL A PROBLEM. COMBUSTION ENGINEERING HAS COMPLETED GAS DISTRIBUTION TESTS AND IS CURRENTLY MODIFYING THE GAS FLOW PATTERN THROUGH THE MODULES. NO RECORD IS KEPT OF THE FGD SYSTEM'S OPERATING HOURS OR SYSTEM AVAILABILITY.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	56
UTILITY NAME	LOUISVILLE GAS & ELECTRIC
UNIT NAME	PADDYS RUN NO. 6
UNIT LOCATION	LOUISVILLE KENTUCKY
UNIT RATING	65 MW
FUEL CHARACTERISTICS	COAL 3.5-4.0 PERCENT SULFOR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIME SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	4/73
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	90 PERCENT RANGE
WATER MAKE UP	33 GAL/LB MOLE SO ₂ REMOVED
SLUDGE DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$57/MW CAPITAL

OPERATIONAL EXPERIENCE FGD SYSTEM OPERATING SATISFACTORILY THE PAST 3 YRS ON LIMESTONE (UNIT OFF ON WELKENS). THESE RUPS ARE MADE FOR AND FUNDED BY COMBUSTION ENGINEERING. AVAILABILITY WAS ESSENTIALLY 100 PERCENT WITH NO PROBLEMS EXPERIENCED IN ANY PART OF THE SYSTEM. THE LIMESTONE SLUDGE IS DEWATERED IN THE EXISTING ROTARY VACUUM FILTERS WITH NO BLINDING OF FILTER CLOTH.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	61
UTILITY NAME	NEVADA POWER
UNIT NAME	REID GARDNER NO 1
UNIT LOCATION	MOAPA NEVADA
UNIT RATING	125 MW
FUEL CHARACTERISTICS	COAL 0.5- 1.0 PERCENT SULFUR
FGD VENDOR	COMBUSTION EQUIP. ASSOCIATES
PROCESS	SODIUM CARBONATE SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	12/73
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO2	
WATER MAKE UP	
SLUDGE DISPOSAL	
UNIT COST	
OPERATIONAL EXPERIENCE	FGD SYSTEM PERFORMANCE IS SATISFACTORILY WITH NO OUTSTANDING MECHANICAL OR PROCESS PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADEQUATE SUPPLY OF THOMA (SODIUM CARBONATE ORE) WHICH RESULTED IN FREQUENT INTERRUPTIONS IN THE OPERATION OF THE FGD SYSTEM. NEVADA POWER IS SUIING THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SODIUM CARBONATE URINE PURCHASED FROM AN ALTERNATE SUPPLIER.

TABLE 3

PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO. 62

UTILITY NAME NEVADA POWER

UNIT NAME RELU GARONER NO 2

UNIT LOCATION MAOPA, NEVADA

UNIT RATING 125 MW

FUEL CHARACTERISTICS COAL 0.5- 1.0 PERCENT SULFUR

FGD VENDOR COMBUSTION EQUIP. ASSOCIATES

PROCESS SODIUM CARBONATE SCRUBBING

NEW OR RETROFIT RETROFIT

START UP DATE 12/73

FGD STATUS OPERATIONAL

EFFICIENCY,
PARTICULATES

SO₂

WATER MAKE UP

SLUDGE DISPOSAL

UNIT COST

OPERATIONAL
EXPERIENCE

FGD SYSTEM PERFORMANCE IS SATISFACTORILY WITH NO OUTSTANDING MECHANICAL OR
PROCESS PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADE-
QUATE SUPPLY OF TRONA (SODIUM CARBONATE ORE) WHICH RESULTED IN FREQUENT
INTERRUPTIONS IN THE OPERATION OF THE FGD SYSTEM. NEVADA POWER IS SUIING
THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SODIUM
CARBONATE BRINE PURCHASED FROM AN ALTERNATE SUPPLIER.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO. 75
UTILITY NAME POTOMAC ELECTRIC & POWER

UNIT NAME DICKERSON NO 3

UNIT LOCATION DICKERSON MARYLAND

UNIT RATING 100 MW

FUEL CHARACTERISTICS COAL 2.0 PERCENT SULFUR

FGD VENDOR CHEMICO

PROCESS MAGNESIUM OXIDE SCRUBBING

NEW OR RETROFIT RETROFIT

START UP DATE 9/73

FGD STATUS OPERATIONAL

EFFICIENCY,
PARTICULATES

SO2

WATER MAKE UP

SLUDGE DISPOSAL

UNIT COST \$707K CAPITAL

OPERATIONAL
EXPERIENCE

THE FGD SYSTEM CONSISTS OF ONE VENTURI STAGE FOR FLY ASH SCRUBBING WITH
WATER FOLLOWED BY A SECOND VENTURI STAGE FOR SO2 SCRUBBING WITH FGD
SLURRY. THE SYSTEM WAS OPERATED FROM OCTOBER 3RD TO THE 14TH. PRESENT-
LY THE SYSTEM IS DOWN TO REPAIR LEAKS IN THE FIRST VENTURI STAGE AS WELL
AS PROBLEMS WITH THE MAGSUS BURNER ELEVATOR. AVAILABILITY SINCE AUGUST
20TH AVERAGED 34 PERCENT.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO. 67
UTILITY NAME SOUTHERN CALIFORNIA EDISON
UNIT NAME KUMAWE NO. 2
UNIT LOCATION LAUGHLIN, NEVADA
UNIT RATING 100 MW
FUEL CHARACTERISTICS COAL 0.5-0.8 PERCENT SULFUR
FGD VENDOR SCE/STERN ROGER
PROCESS LINE SCRUBBING
NEW OR RETROFIT RETROFIT
START UP DATE 11/73
FGD STATUS OPERATIONAL
EFFICIENCY, PARTICULATES
SO₂ 90 PERCENT
WATER MAKE UP 370 GAL/LB MOLE SO₂ REMOVED
SLUDGE DISPOSAL STABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST

OPERATIONAL EXPERIENCE
THE TEST RUNS UNDER THE KUMAWE PLANT OPERATING CONDITIONS HAVE BEEN COMPLETED. THIS EXPERIMENTAL HORIZONTAL MODULE HAS BEEN TEST RUN ON LIMESTONE FOR ABOUT TWO WEEKS IN OCTOBER. THE UNIT IS NOW BACK ON LINE. THE AVERAGE AVAILABILITY OF THIS FGD SYSTEM SINCE ITS INITIAL START UP IN JANUARY 1974 IS ABOUT 80 PERCENT.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO. 93
UTILITY NAME TENNESSEE VALLEY AUTHORITY

UNIT NAME SHAWNEE NO. 10
UNIT LOCATION PADUCAH KENTUCKY

UNIT RATING 30 MW

FUEL CHARACTERISTICS COAL

FGD VENDOR TENNESSEE VALLEY AUTHORITY

PROCESS LIME/LIMESTONE SCRUBBING

NEW OR RETROFIT RETROFIT

START UP DATE 4/72

FGD STATUS OPERATIONAL

EFFICIENCY,
PARTICULATES

SO2

WATER MAKE UP

SLUDGE DISPOSAL

UNIT COST

OPERATIONAL EXPERIENCE THIS IS AN EXPERIMENTAL SYSTEM FUNDED BY USEPA. THERE ARE THREE TYPES OF SCRUBBERS (10 MW EQUIVALENT EACH) BEING EVALUATED UNDER A VARIETY OF OPERATIONAL MODES.

TABLE 4
NUMBER AND TOTAL MW OF FGD SYSTEMS

STATUS	NO. OF UNITS	MW
OPERATIONAL	19	3291.
UNDER CONSTRUCTION	10	6677.
PLANNING		
CONTRACT AWARDED	9	3904.
LETTER OF INTENT	3	530.
REQUESTING/EVALUATING BIDS	13	6667.
CONSIDERING ONLY FGD SYSTEMS	37	16472.
TOTAL	92	37461.

TABLE 5
SUMMARY OF FPD SYSTEMS BY COMPANY

UTILITY	TOTAL				OPERATIONAL				CONSTRUCTION				STATUS				PLANNED			
	NO.	MW	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.	NO.
ALABAMA ELECTRIC COOP	2	450.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2	450.	0
ARIZONA ELECTRIC POWER	2	405.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARIZONA PUBLIC SERVICE	9	516.	1	115.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BASIC ELECTRIC	3	1650.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BOSTON EDISON	4	150.	1	150.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CENTRAL ILLINOIS LIGHT CO.	1	100.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CENTRAL ILLINOIS PUBLIC SERV	1	800.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CINCINNATI GAS AND ELECTRIC	2	1100.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CITY OF KEY WEST	1	97.	1	37.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COLORADO LIC ELECTRIC	4	1352.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COLLINS & SOUTHERN OHIO	2	750.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CONGAMERLTH EDISON	1	167.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DARTMOUTH POWER COOP	1	80.	1	80.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DETROIT EDISON	1	180.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
COLOMBUS LIGHT	2	920.	1	410.	1	510.	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GENERAL MOTORS	1	32.	1	32.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GENERAL PUBLIC UTILITIES	1	650.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ILLINOIS POWER	1	110.	1	110.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
INDIANAPOLIS POWER AND LIGHT	1	515.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KANSAS CITY POWER & LIGHT	3	1000.	3	1000.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KANSAS POWER & LIGHT	4	1825.	2	525.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KENTUCKY UTILITIES	1	84.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LOUISVILLE GAS & ELECTRIC	1	2507.	1	65.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LOUISIANA POWER	4	2120.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LOUISIANA POWER	4	500.	2	250.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NEW ENGLAND ELCC SYSTEM	1	650.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NORTHERN INDIANA POWER SERVICE	3	705.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NORTHERN STATES POWER	2	1550.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PENNSYLVANIA POWER CO.	3	2400.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PHILADELPHIA ELECTRIC	1	120.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
POTOMAC ELECTRIC & POWER	1	100.	1	100.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PUBLIC SERVICE OF INDIANA	1	850.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PUBLIC SERVICE OF NEW JERSEY	4	1715.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S. CAROLINA PUB SERV AUTHORITY	1	140.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SALT RIVER PROJECT	3	2250.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOUTHERN CALIFORNIA EDISON	4	1500.	1	160.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOUTHERN MISSISSIPPI POWER COOP	2	446.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOUTHWESTERN PUBLIC SERVICE	1	343.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SPRINGFIELD UTILITY BOARD	1	200.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TENNESSEE VALLEY AUTHORITY	2	500.	1	30.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TEXAS UTILITIES	5	5972.	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTALS	99	37961.	19	3291.	10	6877.	9	3904.	3	530.	13	6687.	37	16472.						

TABLE B
SUMMARY OF FGD SYSTEMS BY VENDOR

MANUFACTURER/PROCESS	TOTAL		STATUS		CONSTRUCTION	
	NO.	FW	NO.	FW	NO.	FW
AMERICAN AIR FILTER						
LIME SCRUBBING	1	64.	0	0.	1	64.
TOTAL, AMERICAN AIR FILTER	1	64.	0	0.	1	64.
BARBUCK & WILCOX						
LIMESTONE SCRUBBING	2	907.	2	987.	0	0.
TOTAL, BARBUCK & WILCOX	2	907.	2	987.	0	0.
CHEMICO						
LIME SCRUBBING	4	2600.	1	410.	3	2190.
MAGNESIUM OXIDE SCRUBBING	2	250.	2	250.	0	0.
TOTAL, CHEMICO	6	2850.	3	660.	3	2190.
CORROSION ENGINEERING						
LIME SCRUBBING	1	65.	1	65.	0	0.
LIME/LIMESTONE SCRUBBING	1	343.	0	0.	1	343.
LIMESTONE INJECTION A&E SCRUB	4	765.	4	765.	0	0.
LIMESTONE SCRUBBING	2	1360.	0	0.	2	1360.
TOTAL, CORROSION ENGINEERING	8	2533.	5	630.	3	1703.
CORROSION EQUIP. ASSOCIATES						
LIME SCRUBBING	2	720.	0	0.	2	720.
SODIUM CARBONATE SCRUBBING	3	375.	2	250.	1	125.
TOTAL, CORROSION EQUIP. ASSOCIATES	5	1095.	2	250.	3	845.
DAVY POWERGAS						
WELLMAN LORO	1	115.	0	0.	1	115.
TOTAL, DAVY POWERGAS	1	115.	0	0.	1	115.
FOSTER WHEELER						
LIME INJECTION	1	60.	1	60.	0	0.
TOTAL, FOSTER WHEELER	1	60.	1	60.	0	0.
KOCI						
DOUBLE ALKALI	1	32.	1	32.	0	0.
TOTAL, KOCI	1	32.	1	32.	0	0.
MORGANTHO ENVIRO CHEM SYSTEMS						
CATALYTIC OXIDATION	1	110.	1	110.	0	0.
TOTAL, MORGANTHO ENVIRO CHEM SYS	1	110.	1	110.	0	0.

TABLE 6
SUMMARY OF FGD SYSTEMS BY VENDOR

MANUFACTURE/PROCESS	TOTAL		OPERATIONAL		CONSTRUCTION	
	NO.	MG	NO.	MG	NO.	MG
NOT SELECTED						
LIME/LIMESTONE SCRUBBING	1	650.	0	0.	1	650.
TOTAL, NOT SELECTED	1	650.	0	0.	1	650.
PLABODY ENGINEERING						
LIMESTONE SCRUBBING	1	180.	0	0.	1	180.
TOTAL, PLABODY ENGINEERING	1	180.	0	0.	1	180.
RESEARCH COTTELL						
LIMESTONE SCRUBBING	1	115.	1	115.	0	0.
TOTAL, RESEARCH COTTELL	1	115.	1	115.	0	0.
RILEY STICKER/ENGINEERING						
LIMESTONE SCRUBBING	1	100.	0	0.	1	100.
TOTAL, RILEY STICKER/ENGINEERING	1	100.	0	0.	1	100.
SCE/ STERN ROGER						
LIME SCRUBBING	1	160.	1	160.	0	0.
TOTAL, SCE/ STERN ROGER	1	160.	1	160.	0	0.
TENNESSEE VALLEY AUTHORITY						
LIME/LIMESTONE SCRUBBING	1	30.	1	30.	0	0.
LIMESTONE SCRUBBING	1	550.	0	0.	1	550.
TOTAL, TENNESSEE VALLEY AUTHORITY	2	580.	1	30.	1	550.
UNITED ENGINEERS						
MAGNESIUM OXIDE SCRUBBING	1	120.	0	0.	1	120.
TOTAL, UNITED ENGINEERS	1	120.	0	0.	1	120.
UNIVERSAL OIL PRODUCTS						
LIMESTONE SCRUBBING	2	360.	0	0.	2	360.
TOTAL, UNIVERSAL OIL PRODUCTS	2	360.	0	0.	2	360.
ZURN AIR SYSTEMS						
LIMESTONE SCRUBBING	1	37.	1	37.	0	0.
TOTAL, ZURN AIR SYSTEMS	1	37.	1	37.	0	0.
	37	10168.	19	3294.	18	6877.

TABLE 7
SUMMARY OF FGD SYSTEMS BY PROCESS

MW CAPACITY (NO. OF PLANTS)

PROCESS	NO. ON RETROFIT	NO.	1*	NO.	MW	CONSTRUCTION	CONTRACT AMOUNT	LETTER OF INTENT	PLANNED REQUESTING/ CONSIDERING	FOO SYSTEM	TOTAL NO. OF PLANTS
LIME SCRUBBING	N	0	0	4	2400		2 750	0	NO.	MW	NO.
	M	4	715	2	574		2 603	0	NO.	MW	NO.
LIME/LIMESTONE SCRUBBING (PROCESS NOT FOLLOWED)	N	0	0	1	343		0	0	4	1755	3
	M	4	50	1	600		0	0	0	0	1
LIMESTONE SCRUBBING	N	0	1457	4	1600		3 1036	2	405	0	0
	M	5	647	3	890		0	0	0	0	0
SUBTOTAL - LIME/LIMESTONE	N	0	1257	9	4405		5 2086	2	405	12	6237
	M	10	1392	0	2114		2 603	0	0	0	0
CATALYTIC OXIDATION	N	0	0	0	0		0	0	0	0	0
	M	1	110	0	0		0	0	0	0	0
DOUBLE ALKALI	N	0	0	0	0		0	0	0	0	0
	M	1	32	0	0		0	0	0	0	0
MAGNESIUM OXIDE SCRUBBING	N	0	0	0	0		0	0	0	0	0
	M	2	250	1	120		0	0	0	0	0
MAGNESIUM OXIDE SCRUBBING	N	0	0	0	0		0	0	0	0	0
	M	0	0	0	0		0	0	0	0	0
NOT SELECTED	N	0	0	0	0		0	0	0	0	0
	M	0	0	0	0		0	0	0	0	0
SODIUM CARBONATE SCRUBBING	N	0	0	0	0		0	0	0	0	0
	M	2	250	1	120		0	0	0	0	0
WELLMAN LORO	N	0	0	0	0		1 375	0	0	0	0
	M	0	0	1	110		1 340	0	0	0	0
TOTALS	N	0	1257	9	4405		6 4901	3	530	13	6867
	M	10	2034	9	2474		3 943	0	0	0	0
LIME/LIMESTONE % OF TOTAL MW	N	100		100			87	76	91	72	85
	M	80		85			84	0	0	75	75

90

TABLE 3
SUMMARY OF OPERATING FGD SYSTEMS BY
PROCESS AND GENERATING UNITS AS OF 10/74

PROCESS/GENERATING UNITS FGD/NO STARTUP EXPERIENCE/NO

CATALYTIC OXIDATION

WOOD RIVER NO 4	410	10-72	24
	140		24

SCORLE ALKALI

CHEVROLET PARKA 1 2 3 & 4	32	3-74	7
	32		7

LIME INJECTION

ALMA STATION	60	6-71	40
	60		40

LIME SCRUBBING

MURAVE NO 2	160	11-73	11
PAIDYS RUN NO 6	80	4-73	18
PHILLIPS	410	7-73	15
	800		44

LIME/LIMESTONE SCRUBBING

SHAWNEE NO. 10	30	4-72	30
	30		30

LIMESTONE SCRUBBING

MARTIN NO 3	140	11-72	23
MARTIN NO 4	100	3-72	26
LAWRENCE NO 4	420	12-68	70
LAWRENCE NO 5	400	11-71	35
CHOLLA NO 1	115	10-73	12

TABLE A
SUMMARY OF OPERATING FGD SYSTEMS OF
PROCESS AND GENERATING UNITS AS OF 10/74

PROCESS/GENERATING UNITS FGD/AA STARTUP EXPERIENCE(No.)

KEY WEST POWER PLANT 37 10-72 24
LA CYGNE NO 1 020 6-73 16
WILL COUNTY NO 1 167 2-72 32

1904. --- 238

MAGNESION OXIDE SCRUBBING

DICKERSON NO 3 400 9-73 13
MYSTIC NO 6 150 4-72 30

200. --- 43

SODIUM CARBONATE SCRUBBING

REID GARMIN NO 1 125 12-73 10
REID GARMIN NO 2 125 12-73 10

250. --- 20

TABLE 9
SUMMARY OF SLUDGE DISPOSAL PRACTICES FOR OPERATIONAL FGD SYSTEMS AS OF 10/74

PROCESS/GENERATING UNIT

---SLUDGE--- --SLUDGE-- --POND-- --POND--
STABILIZED UNSTABILIZED LINED UNLINED

LINE SCRUBBING

POHAWE NO 2	160			160
PACOYS ROW NO 5		85		85
PHILLIPS	410			410
TOTAL	570.	85.	0.	655.

LIVESTOCK SCRUBBING

PARTRIDGE NO 3	140			140
PARTRIDGE NO 4	100			100
LAURENCE NO 4	120			120
LAURENCE NO 5	400			400
COOPER NO 1	115			115
KEY WEST POWER PLANT				37
LA CYGNE NO 1	820			820
WILL COUNTY NO 1	167		167	
TOTAL	167.	1700.	167.	1737.

TABLE 10
SUMMARY OF FGD WATER LOOP PRACTICES

PARTIAL
OPEN CLOSED CLOSED

CATALYTIC OXIDATION

WOOD RIVER NO 4 0 0 110

0 0 110

DOUBLE ALKALI

CHEVROLET PAKMA 1 2 3 & 4 0 0 32

0 0 32

LINE INJECTION

ALMA STATION 0 0 60

0 0 60

LINE SCRUBBING

MONROE NO 2 0 0 100
PACIFIC RGN NO 6 0 0 65
PHILLIPS 0 410 0

0 410 225

LIME/LIFESTONE SCRUBBING

SHAWNEE NO. 10 0 0 30

0 0 30

LIFESTONE INJECTION & WET SCRUB

HARTMAN NO 3 0 140 0
HARTMAN NO 4 0 100 0
LAWRENCE NO 4 0 145 0
LAWRENCE NO 5 0 400 0

0 765 0

TABLE 10

SUMMARY OF FGD WATER LOOP PRACTICES

PARTIAL
OPEN CLOSED CLOSED

LIMESTONE SCRUBBING

CHOLLA NO 1	115	0	0
KEY WEST POWER PLANT	37	0	0
LA CYGNE NO 1	0	420	0
MILL COUNTY NO 1	0	167	0
	-----	-----	-----
	152	907	0

MAGNESIUM OXIDE SCRUBBING

DICKERSON NO 3	0	0	100
MYSTIC NO 6	0	0	150
	-----	-----	-----
	0	0	250

SODIUM CARBONATE SCRUBBING

REID GARDNER NO 1	125	0	0
REID GARDNER NO 2	125	0	0
	-----	-----	-----
	250	0	0

TABLE 11
OPERATIONAL FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR REINFORCED	SIZE OF FGD UNIT (MW)	PROCESS/VENDOR	FULL CHARACTERISTICS	MO/YR
ARIZONA PUBLIC SERVICE CHOLLA NO. 1	R	115	RESEARCH LOTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	10/73
BOSTON EDISON MYSTIC NO. 6	R	150	CHEMICO MAGNESIUM OXIDE SCRUBBING	OIL 2.5 PERCENT SULFUR	4/72
CITY OF KEY WEST KEY WEST POWER PLANT	N	37	ZURN AIR SYSTEMS LIMESTONE SCRUBBING	OIL 2.4 PERCENT SULFUR	10/72
COMMONWEALTH EDISON WILL COUNTY NO. 1	R	167	HADCOCK & WILCOX LIMESTONE SCRUBBING	COAL 0.6-3.0 PERCENT SULFUR	2/72
DAIRYLAND POWER COOP ALMA STATION	R	80	FOSTER WHEELER LINE INJECTION	COAL 3.0- 3.5 PERCENT SULFUR	6/71
DOUGLASS LIGHT PHILLIPS	R	110	CHEMICO LIME SCRUBBING	COAL 1.0- 2.0 PERCENT SULFUR	7/73
GENERAL MOTORS CHEVROLET PARK 1, 2, 3 & 4	R	32	KUCH DOUBLE ALKALI	COAL 2.5 PERCENT SULFUR	3/74
ILLINOIS POWER WOOD RIVER NO. 4	R	110	NONSANTO LUVIRO CHEM SYSTEMS CATALYTIC OXIDATION	COAL 2.9- 3.2 PERCENT SULFUR	10/72
KANSAS CITY POWER & LIGHT LA CYGNE NO. 1	N	620	HADCOCK & WILCOX LIMESTONE SCRUBBING	COAL 5.2 PERCENT SULFUR	6/73
KANSAS CITY POWER & LIGHT HAWTHORN NO. 4	R	100	COMBUSTION ENGINEERING LIMESTONE INJECTION & ET SCRUB	COAL 0.6- 3.0 PERCENT SULFUR	8/72
KANSAS CITY POWER & LIGHT HAWTHORN NO. 3	R	140	COMBUSTION ENGINEERING LIMESTONE INJECTION & ET SCRUB	COAL 0.6- 3.0 PERCENT SULFUR	11/72
KANSAS POWER & LIGHT LAWRENCE NO. 5	N	400	COMBUSTION ENGINEERING LIMESTONE INJECTION & ET SCRUB	COAL 3.5 PERCENT SULFUR	11/71
KANSAS POWER & LIGHT LAWRENCE NO. 4	R	125	COMBUSTION ENGINEERING LIMESTONE INJECTION & ET SCRUB	COAL 3.5 PERCENT SULFUR	12/68
LOUISVILLE GAS & ELECTRIC PADDOY RUN NO. 6	R	65	COMBUSTION ENGINEERING LIME SCRUBBING	COAL 3.5-4.0 PERCENT SULFUR	4/73
NVADA POWER REID GARDNER NO. 1	R	125	COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING	COAL 0.5- 1.0 PERCENT SULFUR	12/73

920

TABLE 11
OPERATIONAL FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	ALR OR RETRUFLT	SIZE OF FGD UNIT (MW)	PROCESS/VENDOR	FULL CHARACTERISTICS	MO/YR
TEVATA POWER REID GARDNER NO 2	N	125	COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING	COAL 0.5- 1.0 PERCENT SULFUR	12/73
POTOMAC ELECTRIC & POWER DICKERSON NO 3	R	100	CHEMICO MAGNESIUM OXIDE SCRUBBING	COAL 2.0 PERCENT SULFUR	9/73
SOUTHERN CALIFORNIA EDISON ROMAVER NO 2	N	160	SCE/ STERN ROGER LIME SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	11/73
TENNESSEE VALLEY AUTHORITY SHARPE NO. 10	R	30	TENNESSEE VALLEY AUTHORITY LIME/LIMESTONE SCRUBBING	COAL	4/72

TABLE 12
FGD SYSTEMS UNDER CONSTRUCTION AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	PROCESS/VENDOR	FUEL CHARACTERISTICS	MO/YR
CENTRAL ILLINOIS LIGHT CO. DUCK CREEK NO. 1	N	100	RILEY STOKLEY/ENVIRONMENTAL LIMESTONE SCRUBBING	COAL 2.5-3.0 PERCENT SULFUR	3/76
DETROIT EDISON ST. CLAIN NO. 6	R	180	PEABODY ENGINEERING LIMESTONE SCRUBBING	COAL 3.7 PERCENT SULFUR	11/74
SUGGESTHE LIGHT ELMIRA	R	510	CHEMICO LINE SCRUBBING	COAL	5/75
KENTUCKY UTILITIES GREEN RIVER 1 2 & 3	R	64	AMERICAN AIR FILTER LINE SCRUBBING	COAL 3.8 PERCENT SULFUR	4/75
MONTANA POWER COLSTRIP NO. 2	N	360	COMBUSTION EQUIP. ASSOCIATES LINE SCRUBBING	COAL 0.8 PERCENT SULFUR	5/76
MONTANA POWER COLSTRIP NO. 1	N	360	COMBUSTION EQUIP. ASSOCIATES LINE SCRUBBING	COAL 0.8 PERCENT SULFUR	5/75
NEVADA POWER REID GARDNER NO. 3	R	125	COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING	COAL 0.5-1.0 PERCENT SULFUR	6/75
NORTHERN INDIANA PUB SERVICE MITCHELL NO. 11	R	115	DAVY POWERGAS ALLKMAN LUND	COAL 3.2-3.5 PERCENT SULFUR	6/75
NORTHERN STATES POWER SHERBURNE NO. 1	N	600	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL 1.0 PERCENT SULFUR	5/76
NORTHERN STATES POWER SHERBURNE NO. 2	N	600	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL 1.0 PERCENT SULFUR	3/77
PENNSYLVANIA POWER CO. BRUCE MAUSFIELD NO. 2	N	600	CHEMICO LINE SCRUBBING	COAL 4.3 PERCENT SULFUR	6/75
PENNSYLVANIA POWER CO. BRUCE MAUSFIELD NO. 1	N	600	CHEMICO LINE SCRUBBING	COAL 4.3 PERCENT SULFUR	6/75
PHILADELPHIA ELECTRIC EDDYSTONE NO. 1	R	120	UNITED ENGINEERS MAGNESIUM OXIDE SCRUBBING	COAL 2.5 PERCENT SULFUR	12/74
PUBLIC SERVICE OF INDIANA GIBSON NO. 2	R	650	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 1.5 PERCENT SULFUR	5/75
SOUTHERN CALIFORNIA EDISON MORAVE NO. 1	R	160	UNIVERSAL CIL PRODUCTS LIMESTONE SCRUBBING	COAL 0.5-0.8 PERCENT SULFUR	12/74

TABLE 12

FGD SYSTEMS UNDER CONSTRUCTION AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	PROCESS/VENDOR LIME/LIMESTONE SCRUBBING	FUEL CHARACTERISTICS COAL	MO/YR
SOUTHWESTERN PUBLIC SERVICE HARRINGTON, CO 1	N	345	COMBUSTION ENGINEERING LIME/LIMESTONE SCRUBBING	COAL 0.5 PERCENT SULFUR	8/76
SPRINGFIELD UTILITY BOARD SOUTHWEST NO 1	N	200	UNIVERSAL OIL PRODUCTS LIMESTONE SCRUBBING	COAL	6/76
TENNESSEE VALLEY AUTHORITY WIDOWS CREEK NO 8	N	550	TENNESSEE VALLEY AUTHORITY LIMESTONE SCRUBBING	COAL 3.7 PERCENT SULFUR	1/77

TABLE 13

PLANT FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	TYPE ON RETROFIT	SIZE OF FGD UNIT (MW)	VENDOR/PROCESS	FUEL CHARACTERISTICS	M\$/YR
----------------------------------	---------------------	--------------------------	----------------	----------------------	--------

CONTRACTS AWARDED

LOUISVILLE GAS & ELECTRIC CAME RUN NO 4	N	178	AMERICAN AIR FILTER LIME SCRUBBING	COAL 3.5-4.05 PERCENT SULFUR	6/75
COLUMBUS & SOUTHERN OHIO COLUMSVILLE NO 5	N	375	UNIVERSAL OIL PRODUCTS LIME SCRUBBING	COAL	1/76
COLUMBUS & SOUTHERN OHIO CONESVILLE NO 6	N	375	UNIVERSAL OIL PRODUCTS LIME SCRUBBING	COAL	1/78
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 3	N	425	AMERICAN AIR FILTER LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/77
ARIZONA PUBLIC SERVICE CHOLLA NO 2	N	250	RESEARCH CUTTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	6/77
TEXAS UTILITIES MARTIN LAKE NO. 2	N	793	RESEARCH CUTTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	6/77
TEXAS UTILITIES MARTIN LAKE NO. 1	N	793	RESEARCH CUTTRELL LIMESTONE SCRUBBING	COAL 0.4 PERCENT SULFUR	9/76
PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 2	N	340	NAVY POWERGAS KELLMAN LORO	COAL 0.8 PERCENT SULFUR	1/77
PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO 1	N	375	NAVY POWERGAS KELLMAN LORO	COAL 0.8 PERCENT SULFUR	10/77

LETTER OF INTENT SIGNED

NEVADA POWER REID GARNER NO 4	N	125	COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING	COAL 0.5- 1.0 PERCENT SULFUR	6/78
ARIZONA ELECTRIC POWER APACHE NO 3	N	205	RESEARCH CUTTRELL LIMESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	6/79
ARIZONA ELECTRIC POWER APACHE NO 2	N	200	RESEARCH CUTTRELL LIMESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	6/79

REQUESTING/EVALUATING BIOS

CINCINNATI GAS AND ELECTRIC MIAMI FORT NO 8	N	500	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 3.2 PERCENT SULFUR	1/77
KANSAS POWER & LIGHT JEFFERY NO. 1	N	700	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL	6/78
GENERAL PUBLIC UTILITIES	N	650	NOT SELECTED	COAL 2.8 PERCENT SULFUR	4/77

TABLE 15
PLANT FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	VENDOR/PROCESS	FUEL CHARACTERISTICS	MO/YR
ARIZONA PUBLIC SERVICE CHOLLA NO. 3	N	250	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	6/78
CENTRAL ILLINOIS PUBLIC SERV HEWTON NO. 1	N	600	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 2.0-3.2 PERCENT SULFUR	3/77
INDIANAPOLIS POWER AND LIGHT PETERSBURG NO. 3	N	515	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 3.0-3.5 PERCENT SULFUR	4/77
KANSAS POWER & LIGHT JEFFERY NO. 2	N	700	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL	6/79
SOUTHERN MISSISSIPPI PWR COOP HATTIESBURG NO. 2	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
S. CAROLINA PUB SERV AUTHORITY GEORGETOWN NO. 2	N	140	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 1.0 PERCENT SULFUR	5/77
TEXAS UTILITIES MARTIN LAKE NO. 4	N	793	RESEARCH COTTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	12/79
TEXAS UTILITIES MARTIN LAKE NO. 3	N	793	RESEARCH COTTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	12/78
SOUTHERN MISSISSIPPI PWR COOP HATTIESBURG NO. 1	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
TEXAS UTILITIES MONTICELLO NO. 3	N	600	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.8- 1.0 PERCENT SULFUR	12/78

CONSIDERING FGD SYSTEM

LOUISVILLE GAS & ELECTRIC CAME RUN NO. 1	R	110	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/80
COLORADO UTE ELECTRIC HAYDEN NO. 2	R	250	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
COLORADO UTE ELECTRIC CRAIG STATION NO. 1	N	450	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
COLORADO UTE ELECTRIC CRAIG STATION NO. 2	N	450	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
CINCINNATI GAS AND ELECTRIC EAST BEND NO. 1	N	600	NOT SELECTED NOT SELECTED	COAL 0.5- 0.8 PERCENT SULFUR	6/79

TABLE 13
PLANNED FGD SYSTEMS AS OF 10/7/4

UTILITY COMPANY POWER STATION	NEW OR REINFORC	SIZE OF FGD UNIT (MW)	VEHICLE/PROCESS	FUEL CHARACTERISTICS	MO/YR
ALABAMA ELECTRIC COOP BIGWEE NO 3	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.6- 1.5 PERCENT SULFUR	1/79
BASIN ELECTRIC MISSOURI BASIN NO 2	N	550	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/80
BASIN ELECTRIC MISSOURI BASIN NO 3	N	550	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/83
BASIN ELECTRIC MISSOURI BASIN NO 1	N	550	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 4	N	425	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/79
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 1	N	330	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/79
LOUISVILLE GAS & ELECTRIC CANE RUN NO 6	N	277	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/77
LOUISVILLE GAS & ELECTRIC CANE RUN NO 5	N	165	NOT SELECTED LIME SCRUBBING	COAL 3.5-4.05 PERCENT SULFUR	6/76
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 2	N	330	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/76
LOUISVILLE GAS & ELECTRIC CANE RUN NO 2	N	107	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/80
ARIZONA PUBLIC SERVICE CHOLLA NO 4	N	372	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
PENNSYLVANIA POWER CO. BRUCE MANSFIELD NO. 3	N	800	NOT SELECTED LIME SCRUBBING	COAL 4.3 PERCENT SULFUR	6/78
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 2	N	175	CHEMICO NOT SELECTED	COAL 0.6 PERCENT SULFUR	12/76
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 3	N	229	CHEMICO LIME SCRUBBING	COAL 0.8 PERCENT SULFUR	3/77
NORTHERN INDIANA PUB SERVICE BAILY NO. 7	N	190	NOT SELECTED NOT SELECTED	COAL 3 PERCENT SULFUR	0/ 0

TABLE 13
PLANNED FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	AGE OR RETOFIT	SIZE OF FGD UNIT (MW)	VENDOR/PROCESS	FUEL CHARACTERISTICS	MO/YR
GORTHERN INDIANA PUB SERVICE DAILY NO. 6	K	400	NOT SELECTED NOT SELECTED	COAL 3 PERCENT SULFUR	0/0
ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 4	R	800	CHEMICO LIME SCRUBBING	COAL 0.6 PERCENT SULFUR	4/77
COLORADO UTE ELECTRIC HAYDEN NO. 1	R	202	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
NEW ENGLAND ELEC SYSTEM BRAYTON POINT NO. 3	R	650	NOT SELECTED NOT SELECTED	COAL 0.3 PERCENT SULFUR	0/0
ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 1	K	175	CHEMICO LIME SCRUBBING	COAL 0.6 PERCENT SULFUR	10/76
MONTANA POWER COLSTRIP NO. 3	N	700	NOT SELECTED NOT SELECTED	COAL 0.7 PERCENT SULFUR	7/78
LOUISVILLE GAS & ELECTRIC CAKE RUN NO. 3	R	137	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/80
MONTANA POWER COLSTRIP NO. 4	N	700	NOT SELECTED NOT SELECTED	COAL 0.7 PERCENT SULFUR	0/79
ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 5	R	800	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 0.6 PERCENT SULFUR	6/77
ALABAMA ELECTRIC COOP DINGLE NO. 2	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.6- 1.5 PERCENT SULFUR	1/79
PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 3	N	500	NOT SELECTED NOT SELECTED	COAL 0.8 PERCENT SULFUR	5/78
PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 4		500	NOT SELECTED NOT SELECTED	COAL 0.6 PERCENT SULFUR	5/80
SALT RIVER PROJECT NAVAJO NO. 1	N	750	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	3/76
SOUTHERN CALIFORNIA Edison MONTE NO. 1	R	640	NOT SELECTED LIME SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	6/77
SALT RIVER PROJECT NAVAJO NO. 3	N	750	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	3/77

TABLE 13
PLANNED FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	VENDOR/PROCESS	FUEL CHARACTERISTICS	MO/YR
SALT RIVER PROJECT NAVAJO NO 2	"	750	NOT SELECTED LIME/LINESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	10/76
SOUTHERN CALIFORNIA Edison MORRIS NO 2	R	640	NOT SELECTED LINE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	6/77

APPENDIX A
FLUE GAS DESULFURIZATION SYSTEM STATUS REPORT
OCTOBER 1974

TABLE 1
SUMMARY OF FGD SYSTEMS

FGD STATUS REPORT 10/74

I.D. NUMBER AND COMPANY NAME	UNIT NAME	LOCATION	START UP DATE	STATUS
1 ALABAMA ELECTRIC COOP	BILOXI NO 2	JACKSON ALABAMA	1-79	6
2 ALABAMA ELECTRIC COOP	BILOXI NO 3	JACKSON ALABAMA	1-79	6
3 ARIZONA ELECTRIC POWER	APACHE NO 2	COCOA ARIZONA	6-79	4
4 ARIZONA ELECTRIC POWER	APACHE NO 3	COCOA ARIZONA	6-79	4
5 ARIZONA PUBLIC SERVICE	CHOLLA NO 1	JOSEPH CITY ARIZONA	10-73	1
6 ARIZONA PUBLIC SERVICE	CHOLLA NO 2	JOSEPH CITY ARIZONA	6-77	3
7 ARIZONA PUBLIC SERVICE	CHOLLA NO 3	JOSEPH CITY ARIZONA	6-78	3
8 ARIZONA PUBLIC SERVICE	CHOLLA NO 4	JOSEPH CITY ARIZONA	6-73	6
9 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 1	FARMINGTON NEW MEXICO	10-76	6
10 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 2	FARMINGTON NEW MEXICO	12-76	6
11 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 3	FARMINGTON NEW MEXICO	3-77	6
12 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 4	FARMINGTON NEW MEXICO	4-77	6
13 ARIZONA PUBLIC SERVICE	FOUR CORNERS NO 5	FARMINGTON NEW MEXICO	6-77	6
14 BASIN ELECTRIC	MISSOURI BASIN NO 1	WHEATLAND WYOMING	6-79	6
15 BASIN ELECTRIC	MISSOURI BASIN NO 2	WHEATLAND WYOMING	6-80	6
16 BASIN ELECTRIC	MISSOURI BASIN NO 3	WHEATLAND WYOMING	6-83	6
17 BOSTON Edison	MYSTIC NO 6	EVERETT MASSACHUSETTS	4-72	1
18 CENTRAL ILL. ELECTRIC CO.	DUCK CREEK NO.1	CANTON ILLINOIS	3-76	2
19 CENTRAL ILL. ELECTRIC CO.	ALTON NO.1	ALTON ILLINOIS	3-77	3
20 VINCINNATI GAS AND ELECTRIC	EAST RENO NO 1	RAIDIT NASH KENTUCKY	6-79	6
21 VINCINNATI GAS AND ELECTRIC	MIAMI FORT NO 6	NORTH BEND OHIO	1-77	1
22 CITY OF PEL-AST	ALY WEST POWER PLANT	ALY WEST FLORIDA	10-72	1
23 COLORADO UTE ELECTRIC	CHAIK STATION NO.1	CHAIK COLORADO	6-78	6
24 COLORADO UTE ELECTRIC	CHAIK STATION NO.2	CHAIK COLORADO	6-78	6
25 COLORADO UTE ELECTRIC	MAIDEN NO 1	MAIDEN COLORADO	6-78	6
26 COLORADO UTE ELECTRIC	MAIDEN NO 2	MAIDEN COLORADO	6-78	6
27 COLORADO UTE ELECTRIC	COLESVILLE NO 5	COLESVILLE OHIO	1-76	3
28 COLORADO UTE ELECTRIC	COLESVILLE NO 6	COLESVILLE OHIO	1-76	3
29 COMINGWORTH Edison	WILL COUNTY NO 1	ROCKFORD ILLINOIS	2-72	1
30 DALLAS AND POWER COOP	ALMA STATION	ALMA WISCONSIN	6-71	1
31 DETROIT Edison	ST. CLAIR NO 6	HELL RIVER MICHIGAN	11-74	2
32 LOUISIANA ELECTRIC	LEONA	CLARKA PENNSYLVANIA	5-73	2
33 LOUISIANA ELECTRIC	PHILLIPS	SOUTH PLIGHT PENNSYLVANIA	7-73	1
34 FEDERAL Edison	CLEVELAND PARKS 1, 2, 3 & 4	PARTA OHIO	3-76	1
35 FEDERAL PUBLIC UTILITIES	MOORE CITY NO.3	MOORE CITY PENNSYLVANIA	4-77	5
36 ILLINOIS POWER	MOORE RIVER NO 4	EAST ALTON ILLINOIS	10-72	1
37 INDIANAPOLIS POWER AND LIGHT	PETERSBURG NO 3	PETERSBURG INDIANA	4-77	5
38 KANSAS CITY POWER & LIGHT	MAATHORN NO 3	KANSAS CITY MISSOURI	11-72	1
39 KANSAS CITY POWER & LIGHT	MAATHORN NO 4	KANSAS CITY MISSOURI	8-72	1
40 KANSAS CITY POWER & LIGHT	LA CYGNE NO 1	LA CYGNE KANSAS	6-73	1
41 KANSAS POWER & LIGHT	JEFFERY NO. 1	ST MARY KANSAS	6-78	5
42 KANSAS POWER & LIGHT	JEFFERY NO. 2	ST MARY KANSAS	6-79	5

1. OPERATIONAL UNITS
2. UNITS UNDER CONSTRUCTION
3. PLANNED - CONTRACT AWARDED
4. PLANNED - LETTER OF INTENT SIGNED
5. PLANNED - REQUESTING/EVALUATING BIDS
6. CONSIDERING ONLY FGD SYSTEMS

TABLE 1
SUMMARY OF FGD SYSTEMS

FGD STATUS REPORT 11/74

PLANT NUMBER AND COMPANY NAME	UNIT NAME	LOCATION	START UP DATE	STATUS
43 KANSAS POWER & LIGHT	LAURENCE NO 4	LAURENCE KANSAS	12-68	1
44 KANSAS POWER & LIGHT	LAURENCE NO 5	LAURENCE KANSAS	11-71	1
45 KENTUCKY UTILITIES	OREN RIVER 1, 2 & 3	SOUTH CARLTON KENTUCKY	4-75	2
46 LOUISVILLE GAS & ELECTRIC	CARL RUN NO 1	LOUISVILLE KENTUCKY	6-80	6
47 LOUISVILLE GAS & ELECTRIC	CARL RUN NO 2	LOUISVILLE KENTUCKY	6-80	6
48 LOUISVILLE GAS & ELECTRIC	CARL RUN NO 3	LOUISVILLE KENTUCKY	6-80	6
49 LOUISVILLE GAS & ELECTRIC	CARL RUN NO 4	LOUISVILLE KENTUCKY	6-75	3
50 LOUISVILLE GAS & ELECTRIC	CARL RUN NO 5	LOUISVILLE KENTUCKY	6-76	6
51 LOUISVILLE GAS & ELECTRIC	CARL RUN NO 6	LOUISVILLE KENTUCKY	6-77	6
52 LOUISVILLE GAS & ELECTRIC	HALL CREEK NO 1	LOUISVILLE KENTUCKY	6-79	6
53 LOUISVILLE GAS & ELECTRIC	HALL CREEK NO 2	LOUISVILLE KENTUCKY	6-78	6
54 LOUISVILLE GAS & ELECTRIC	HALL CREEK NO 3	LOUISVILLE KENTUCKY	6-77	3
55 LOUISVILLE GAS & ELECTRIC	HALL CREEK NO 4	LOUISVILLE KENTUCKY	6-79	6
56 LOUISVILLE GAS & ELECTRIC	PAIDERS RUN NO 4	LOUISVILLE KENTUCKY	4-73	1
57 INDIANA POWER	COLSTRIP NO 1	COLSTRIP MONTANA	5-75	2
58 MONTANA POWER	COLSTRIP NO 2	COLSTRIP MONTANA	5-76	2
59 MONTANA POWER	COLSTRIP NO 3	COLSTRIP MONTANA	7-78	6
60 MONTANA POWER	COLSTRIP NO 4	COLSTRIP MONTANA	0-79	6
61 FLAVIA POWER	RELIO GARDNER NO 1	MAOPA NEVADA	12-73	1
62 FLAVIA POWER	RELIO GARDNER NO 2	MAOPA NEVADA	12-73	1
63 FLAVIA POWER	RELIO GARDNER NO 3	MAOPA NEVADA	6-75	2
64 FLAVIA POWER	RELIO GARDNER NO 4	MAOPA NEVADA	6-78	4
65 NEW JERSEY ELECTRIC SYSTEM	UNION POINT NO. 3	SOMERSET MASSACHUSETTS	0-0	6
66 PORTLAND INDIANA PWS SERVICE	DAILY NO. 7	CHESTERTON INDIANA	0-0	6
67 PORTLAND INDIANA PWS SERVICE	DAILY NO. 8	CHESTERTON INDIANA	0-0	6
68 PORTLAND INDIANA PWS SERVICE	PITCHELL NO 11	GARY INDIANA	6-75	2
69 PORTLAND INDIANA PWS SERVICE	SOCORRO NO 1	ELKEM MINNESOTA	5-76	2
70 PORTLAND INDIANA PWS SERVICE	SOCORRO NO 2	ELKEM MINNESOTA	3-77	2
71 PENNSYLVANIA POWER CO.	BRUCE MARSHFIELD NO. 1	SHIPPINGPORT PENNSYLVANIA	6-75	2
72 PENNSYLVANIA POWER CO.	BRUCE MARSHFIELD NO. 2	SHIPPINGPORT PENNSYLVANIA	6-75	2
73 PENNSYLVANIA POWER CO.	BRUCE MARSHFIELD NO. 3	SHIPPINGPORT PENNSYLVANIA	8-78	6
74 PHILADELPHIA ELECTRIC	EDVISTONE NO 1	EDVISTONE PENNSYLVANIA	12-74	2
75 FORT-AC ELECTRIC & POWER	DICKERSON NO 3	DICKERSON MARYLAND	9-73	1
76 PUBLIC SERVICE OF INDIANA	GLASSBORO NO. 2	PRINCETON INDIANA	5-75	2
77 PUBLIC SERVICE OF NEW MEXICO	SALT JOAN NO 1	WATERFLOW NEW MEXICO	10-77	3
78 PUBLIC SERVICE OF NEW MEXICO	SALT JOAN NO 2	WATERFLOW NEW MEXICO	1-77	3
79 PUBLIC SERVICE OF NEW MEXICO	SALT JOAN NO. 3	WATERFLOW NEW MEXICO	5-78	6
80 PUBLIC SERVICE OF NEW MEXICO	SALT JOAN NO. 4	WATERFLOW NEW MEXICO	5-80	6
81 S. CAROLINA PWS SERV AUTHORITY	GEORGETOWN NO 2	GEORGETOWN SOUTH CAROLINA	5-77	5
82 SALT RIVERS PROJECT	RAVADJO NO 1	PAGE ARIZONA	3-76	6
83 SALT RIVERS PROJECT	RAVADJO NO 2	PAGE ARIZONA	10-76	6
84 SALT RIVERS PROJECT	RAVADJO NO 3	PAGE ARIZONA	3-77	6

1. OPERATIONAL UNITS
2. UNITS UNDER CONSTRUCTION
3. PLANNED - CONTRACT AWARDED
4. PLANNED - LETTER OF INTENT SIGNED
5. PLANNED - REQUESTING/EVALUATING BIDS
6. PLANNED - CONSIDERING ONLY FGD SYSTEMS

TABLE 1
SUMMARY OF FGD SYSTEMS

FGD STATUS REPORT 10/74

I.D. NUMBER AND COMPANY NAME	UNIT NAME	LOCATION	START UP DATE	STATUS
85 SOUTHERN CALIFORNIA Edison	MORAVE NO 1	LAUGHLIN NEVADA	6-77	6
86 SOUTHERN CALIFORNIA Edison	MORAVE NO 1	LAUGHLIN NEVADA	12-74	2
87 SOUTHERN CALIFORNIA Edison	MORAVE NO 2	LAUGHLIN NEVADA	11-73	1
88 SOUTHERN CALIFORNIA Edison	MORAVE NO 2	LAUGHLIN NEVADA	6-77	6
89 SOUTHERN MISSISSIPPI PWR COOP	HATTIESBURG NO.1	HATTIESBURG MISSISSIPPI	6-79	5
90 SOUTHERN MISSISSIPPI PWR COOP	HATTIESBURG NO.2	HATTIESBURG MISSISSIPPI	6-79	5
91 SOUTHERN PUBLIC SERVICE	HARRINGTON NO 1	APARILLO TEXAS	8-76	2
92 SOUTHERN UTILITY BOARD	SOUTHALST NO 1	SPRINGFIELD MISSOURI	6-76	2
93 TENNESSEE VALLEY AUTHORITY	SHAWNEE NO. 10	PAUCAN KENTUCKY	4-72	1
94 TENNESSEE VALLEY AUTHORITY	SHAWNEE LAKE NO. 6	BRIDGEPORT ALABAMA	1-77	2
95 TEXAS UTILITIES	MARTIN LAKE NO. 1	TATUM TEXAS	9-76	3
96 TEXAS UTILITIES	MARTIN LAKE NO.2	TATUM TEXAS	6-77	3
97 TEXAS UTILITIES	MARTIN LAKE NO.3	TATUM TEXAS	12-78	5
98 TEXAS UTILITIES	MARTIN LAKE NO.4	TATUM TEXAS	12-79	5
99 TEXAS UTILITIES	MONTICELLO NO 3	MT PLEASANT TEXAS	12-76	5

1. OPERATIONAL UNITS
2. UNITS UNDER CONSTRUCTION
3. PLANNED - CONTRACT AWARDED
4. PLANNED - LETTER OF INTENT SIGNED
5. PLANNED - REQUESTING/EVALUATING BIDS
6. PLANNED - CONSIDERING ONLY FGD SYSTEMS

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	CURRENT MONTH
POWER STATION	
1.0. JUPITER 1	INSTALLATION OF A FLUE GAS DESULFURIZATION SYSTEM IS BEING CONSIDERED. IF A SYSTEM IS INSTALLED, IT WILL PROBABLY BE THE NET LIMESTONE SCRUBBING PROCESS.
ARIZONA ELECTRIC COOP	
205 50 - 120	
COAL 0.9 - 1.5 PERCENT SULFUR	
NOT SELECTED	
LIMESTONE SCRUBBING	
STARTUP 1/75	
1.7. JUPITER 2	INSTALLATION OF A FLUE GAS DESULFURIZATION SYSTEM IS BEING CONSIDERED. IF A SYSTEM IS INSTALLED, IT WILL PROBABLY BE THE NET LIMESTONE SCRUBBING PROCESS.
ARIZONA ELECTRIC COOP	
205 50 - 120	
COAL 0.9 - 1.5 PERCENT SULFUR	
NOT SELECTED	
LIMESTONE SCRUBBING	
STARTUP 1/75	
1.8. JUPITER 3	
ARIZONA ELECTRIC COOP	
205 50 - 120	
COAL 0.9 - 1.5 PERCENT SULFUR	
NOT SELECTED	
LIMESTONE SCRUBBING	
STARTUP 1/75	
1.9. JUPITER 4	LETTER OF INTENT TO INSTALL AN FGD SYSTEM WAS ISSUED TO RESEARCH COTTRELL.
ARIZONA ELECTRIC COOP	
205 50 - 120	
COAL 0.9 - 1.5 PERCENT SULFUR	
NOT SELECTED	
LIMESTONE SCRUBBING	
STARTUP 1/75	
1.10. JUPITER 5	LETTER OF INTENT TO INSTALL AN FGD SYSTEM WAS ISSUED TO RESEARCH COTTRELL.
ARIZONA ELECTRIC COOP	
205 50 - 120	
COAL 0.9 - 1.5 PERCENT SULFUR	
NOT SELECTED	
LIMESTONE SCRUBBING	
STARTUP 1/75	
1.11. JUPITER 6	BOTH MODULES WERE IN SERVICE DURING THE PAST 30 DAYS, EXCEPT FOR ONE DAY PER MODULE TO CORRECT A VIBRATION PROBLEM IN THE REHEATER UNITS. THE INSTALLATION OF CARTRIDGE PILES IN THE INLET DUCT TO EACH REHEATER TO Dampen THE VIBRATION IS CONSIDERED A PARTIAL SUCCESS BY APS. FGD SYSTEM AVAILABLE WAS 93 PERCENT.
ARIZONA ELECTRIC COOP	
205 50 - 120	
COAL 0.9 - 1.5 PERCENT SULFUR	
NOT SELECTED	
LIMESTONE SCRUBBING	
STARTUP 10/75	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

CURRENT MONTH

POWER STATION

1.0. UNIT 6
ARIZONA PUBLIC SERVICE
CHOLLA NO. 2
250 T. - 18%
COAL 0.4% PERCENT SULFUR
RESEARCH COTTEPIL
LIVESTOCK SCRUBBING
STARTUP 8/77

CONTRACT TO INSTALL FGD SYSTEM ON CHOLLA NO. 3 HAS BEEN AWARDED TO
RESEARCH COTTEPIL.

1.0. UNIT 7
ARIZONA PUBLIC SERVICE
CHOLLA NO. 3
250 T. - 18%
COAL 0.4% PERCENT SULFUR
LIVESTOCK SCRUBBING
STARTUP 6/78

ARIZONA PUBLIC SERVICE IS OUT FOR BIDS AND SOME PROPOSALS HAVE BEEN
RECEIVED AND ARE BEING EVALUATED.

1.0. UNIT 8
ARIZONA PUBLIC SERVICE
CHOLLA NO. 4
312 T. - 18%
COAL 0.4% PERCENT SULFUR
LIVESTOCK SCRUBBING
STARTUP 6/78

PLANS FOR SO₂ CONTROL SYSTEM ARE IN THE EARLY STAGE. APS IS CURRENTLY
WORKING ON THE DESIGN OF UNITS 2 AND 3 AND HAS NOT FIRMED UP PLANS FOR
UNIT 4.

1.0. UNIT 9
ARIZONA PUBLIC SERVICE
CHOLLA NO. 1
175 T. - 18%
COAL 0.4% PERCENT SULFUR
LIVESTOCK SCRUBBING
STARTUP 12/76

EXISTING SCRUBBERS REMOVE ABOUT 30 PERCENT OF THE SO₂. THE VENTURIS WILL
BE UPGRADED TO COMPLY WITH APPLICABLE SO₂ REGULATIONS.

1.0. UNIT 10
ARIZONA PUBLIC SERVICE
CHOLLA NO. 2
175 T. - 18%
COAL 0.4% PERCENT SULFUR
LIVESTOCK SCRUBBING
STARTUP 12/76

EXISTING SCRUBBERS REMOVE ABOUT 30 PERCENT OF THE SO₂. THE VENTURIS WILL
BE UPGRADED TO COMPLY WITH APPLICABLE SO₂ REGULATIONS.

TABLE 2
STATUS OF FGD SYSTEMS DURING 1974

COMPANY	POWER STATION	CURRENT MONTH
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 3	I.D. NUMBER 11	EXISTING SCRUBBERS REMOVE ABOUT 30 PERCENT OF THE SO ₂ . THE VENTURIS WILL BE UPGRADED TO COMPLY WITH APPLICABLE SO ₂ REGULATIONS.
	229 1% - METROIT	
	COAL 6.0 PERCENT SULFUR	
	CHENICO LINE SCRUBBING STARTUP 8/77	
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 4	I.D. NUMBER 12	ESPS ARE USED FOR PARTICULATE EMISSION CONTROL. FGD SYSTEM IS STILL IN THE PLANNING STAGE.
	266 1% - METROIT	
	COAL 6.0 PERCENT SULFUR	
	CHENICO LINE SCRUBBING STARTUP 9/77	
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 5	I.D. NUMBER 13	ESPS ARE USED FOR PARTICULATE EMISSION CONTROL. FGD SYSTEM IS STILL IN THE PLANNING STAGE.
	266 1% - METROIT	
	COAL 6.0 PERCENT SULFUR	
	CHENICO LINE SCRUBBING STARTUP 9/77	
MISSOURI ELECTRIC MISSOURI BASIN NO 1	I.D. NUMBER 14	THE COMPANY IS EVALUATING ALTERNATIVE PROCESSES FOR SO ₂ EMISSION CONTROL. LIMESTONE SCRUBBING IS THE MOST LIKELY PROCESS TO BE SELECTED. THE FINAL DECISION WILL BE MADE IN A FEW MONTHS.
	250 1% - NEW	
	COAL	
	NOT SELECTED LIMESTONE SCRUBBING STARTUP 6/79	
MISSOURI ELECTRIC MISSOURI BASIN NO 2	I.D. NUMBER 15	THE COMPANY IS EVALUATING ALTERNATIVE PROCESSES FOR SO ₂ EMISSION CONTROL. LIMESTONE SCRUBBING IS THE MOST LIKELY PROCESS TO BE SELECTED. THE FINAL DECISION WILL BE MADE IN A FEW MONTHS.
	250 1% - NEW	
	COAL	
	NOT SELECTED LIMESTONE SCRUBBING STARTUP 6/79	

STATUS OF FCC SYSTEMS DURING FALL 2

Table 2

STATUS OF FCC SYSTEMS DURING 10/7/74

Page	Initial	Current Month
10. 10/10/76	10	10
11. 11/10/76	11	11
12. 12/10/76	12	12
13. 1/10/77	13	13
14. 2/10/77	14	14
15. 3/10/77	15	15
16. 4/10/77	16	16
17. 5/10/77	17	17
18. 6/10/77	18	18
19. 7/10/77	19	19
20. 8/10/77	20	20
21. 9/10/77	21	21
22. 10/10/77	22	22
23. 11/10/77	23	23
24. 12/10/77	24	24
25. 1/10/78	25	25
26. 2/10/78	26	26
27. 3/10/78	27	27
28. 4/10/78	28	28
29. 5/10/78	29	29
30. 6/10/78	30	30
31. 7/10/78	31	31
32. 8/10/78	32	32
33. 9/10/78	33	33
34. 10/10/78	34	34
35. 11/10/78	35	35
36. 12/10/78	36	36
37. 1/10/79	37	37
38. 2/10/79	38	38
39. 3/10/79	39	39
40. 4/10/79	40	40
41. 5/10/79	41	41
42. 6/10/79	42	42
43. 7/10/79	43	43
44. 8/10/79	44	44
45. 9/10/79	45	45
46. 10/10/79	46	46
47. 11/10/79	47	47
48. 12/10/79	48	48
49. 1/10/80	49	49
50. 2/10/80	50	50
51. 3/10/80	51	51
52. 4/10/80	52	52
53. 5/10/80	53	53
54. 6/10/80	54	54
55. 7/10/80	55	55
56. 8/10/80	56	56
57. 9/10/80	57	57
58. 10/10/80	58	58
59. 11/10/80	59	59
60. 12/10/80	60	60
61. 1/10/81	61	61
62. 2/10/81	62	62
63. 3/10/81	63	63
64. 4/10/81	64	64
65. 5/10/81	65	65
66. 6/10/81	66	66
67. 7/10/81	67	67
68. 8/10/81	68	68
69. 9/10/81	69	69
70. 10/10/81	70	70
71. 11/10/81	71	71
72. 12/10/81	72	72
73. 1/10/82	73	73
74. 2/10/82	74	74
75. 3/10/82	75	75
76. 4/10/82	76	76
77. 5/10/82	77	77
78. 6/10/82	78	78
79. 7/10/82	79	79
80. 8/10/82	80	80
81. 9/10/82	81	81
82. 10/10/82	82	82
83. 11/10/82	83	83
84. 12/10/82	84	84
85. 1/10/83	85	85
86. 2/10/83	86	86
87. 3/10/83	87	87
88. 4/10/83	88	88
89. 5/10/83	89	89
90. 6/10/83	90	90
91. 7/10/83	91	91
92. 8/10/83	92	92
93. 9/10/83	93	93
94. 10/10/83	94	94
95. 11/10/83	95	95
96. 12/10/83	96	96
97. 1/10/84	97	97
98. 2/10/84	98	98
99. 3/10/84	99	99
100. 4/10/84	100	100

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
I.D. NUMBER 21 CINCINNATI GAS AND ELECTRIC MIAMI FORT NG 4	500 KW - NEW COAL 3.2 PERCENT SULFUR NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 1/77	THE COMPANY IS OBTAINING BIDS FOR AN FGD SYSTEM.
I.D. NUMBER 22 CITY OF KEY WEST KEY WEST POWER PLANT	37 KW - EX COAL 2.8 PERCENT SULFUR LIME/LIMESTONE SCRUBBING STARTUP 10/72	PLANT HAS BEEN SHUT DOWN SINCE SEPT 19TH BECAUSE OF A BOILER 1 D FAN MOTOR FAILURE. THEY HOPE TO BE OPERATIONAL THE WEEK OF OCTOBER 21ST. MECHANICAL FAILURES IN THE FGD SYSTEM WERE FIXED-FGD OPERATED ABOUT A WEEK AND A HALF BEFORE THE PRESENT BOILER SHUTDOWN. DURING SHUTDOWN THE EXHAUSTION JOINTS ON THE DUCTWORK TO THE SCRUBBER WERE REPLACED FOR THE 2ND TIME. SOME PLUGGING REMOVED AT SHUTDOWN.
I.D. NUMBER 23 COLORADO UTL ELECTRIC COLORADO STATE COAL	450 KW - EX COAL 6.45 PERCENT SULFUR NOT SELECTED LIMESTONE SCRUBBING STARTUP 6/77	AS WITH PLANTS 1 & 2 BOTH LIMESTONE AND SODIUM CARBONATE SYSTEMS ARE BEING EVALUATED. USING THE ANTICIPATED 0.45 PERCENT SULFUR COAL AN FGD SYSTEM WITH APPROX. 70 PERCENT EFF. WILL BE REQUIRED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. PRESENT EVALUATION FAVORS A LIMESTONE SCRUBBER WITH SLUDGE DISPOSAL IN A MINE SITE.
I.D. NUMBER 24 COLORADO UTL ELECTRIC COLORADO STATE COAL	450 KW - EX COAL 6.45 PERCENT SULFUR NOT SELECTED LIMESTONE SCRUBBING STARTUP 6/77	AS WITH PLANTS 1 & 2 BOTH LIMESTONE AND SODIUM CARBONATE SYSTEMS ARE BEING EVALUATED. USING THE ANTICIPATED 0.45 PERCENT SULFUR COAL AN FGD SYSTEM WITH APPROX. 70 PERCENT EFF. WILL BE REQUIRED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. PRESENT EVALUATION SEEMS TO FAVOR A LIMESTONE SCRUBBER WITH SLUDGE DISPOSAL IN A MINE SITE.
I.D. NUMBER 25 COLORADO UTL ELECTRIC MAYOR NG 1	202 KW - RETROFIT COAL 6.45 PERCENT SULFUR NOT SELECTED LIMESTONE SCRUBBING STARTUP 8/79	COLORADO UTL IS PROCEEDING WITH ENGINEERING EVALUATION OF BOTH LIME-STONE & SODIUM CARBONATE SYSTEMS. USING THE PRESENT 0.45 PERCENT SULFUR COAL AN FGD UNIT WITH APPROX. 70 PERCENT EFF. WOULD BE NEEDED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. IT IS PROBABLE THAT A LIMESTONE SCRUBBER WILL BE CHOSEN WITH DISPOSAL IN A MINE SITE.

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
COLORADO JIE ELECTRIC	I.D. NUMBER 26	COLORADO JIE IS PROCEEDING WITH ENGINEERING EVALUATION OF BOTH LIMC-STAGE & SODIUM CARBONATE SYSTEMS. USING THE PRESENT 0.45 PERCENT SULFUR COAL AT FGD UNIT WITH APPROX. 70 PERCENT EFF. WOULD BE NEEDED TO MEET THE 1978 COLORADO SO ₂ EMISSION REGS OF 150PPM. IT IS PROBABLY THAT A Limestone SCRUBBER WILL BE CHOSEN WITH DISPOSAL IN A MINE SITE.
	250 PA - RETROFIT	
	COAL 0.45 PERCENT SULFUR	
	NOT SELECTED	
LIMESTONE SCRUBBING	STARTUP 6/76	
UNIVERSAL OIL PRODUCTS	I.D. NUMBER 27	U.O.P. IS PRESENTLY IN THE ENGINEERING DESIGN AND PROCUREMENT STAGE.
	COLUMBUS & SOUTHERN OHIO	
	COLUMBUSVILLE 10 5	CONSTRUCTION IS SCHEDULED TO START MARCH 1975.
	375 PA - NEW	
LIMESTONE SCRUBBING	STARTUP 1/76	
UNIVERSAL OIL PRODUCTS	I.D. NUMBER 26	U.O.P. HAS BEEN DELAYED FOR ABOUT A YEAR. CONSTRUCTION IS NOW SCHEDULED TO START JANUARY 1977. UNIT NOW SCHEDULED FOR COMPLETION IN JANUARY 1978
	COLUMBUS & SOUTHERN OHIO	
	COLUMBUSVILLE 10 6	
	375 PA - NEW	
LIMESTONE SCRUBBING	STARTUP 1/76	
UNIVERSAL OIL PRODUCTS	I.D. NUMBER 29	MODULE A WAS DOWN 3 TIMES LAST MONTH ONCE TO CLEAN DEPOSITS FROM THE VENTURI INLET AND TWICE DUE TO BOILER RELATED PROBLEMS. DEMISTER CONTINUED TO DETERIORATE SLIGHTLY AND WILL HAVE TO BE HAND WASHED DURING SCHEDULED BOILER OUTAGE. OPERATION MOST OF THE MONTH WAS WITH A MEDIUM SULFUR COAL BLEND. AVAILABILITY WAS 85 PERCENT. MODULE B IS STILL DOWN.
	WILL COUNTY 10 1	
	167 PA - RETROFIT	
	COAL 0.6-0.8 PERCENT SULFUR	
LIMESTONE SCRUBBING	STARTUP 2/72	
UNIVERSAL OIL PRODUCTS	I.D. NUMBER 30	EXPERIMENTAL WET LIME FURNACE INJECTION SYSTEM WITH NO TAIL END SCRUBBING IS USED. OUTSTANDING PROBLEMS ARE FURNACE TUBE PLUGGAGE WHICH NECESSITATES FREQUENT CLEANUP AND SHUT OUT LIME GRINDING AND INJECTION EQUIPMENT.
	WILL COUNTY 10 2	
	167 PA - RETROFIT	REMOVAL EFFICIENCY WAS FROM 25 TO 40 PERCENT. THE FGD SYSTEM DID NOT RUN MORE THAN TWO DAYS AT A TIME. AVAILABILITY SINCE STARTUP NEVER EXCEEDED 25 PERCENT.
	COAL 5.0- 5.5 PERCENT SULFUR	
LIMESTONE SCRUBBING	STARTUP 6/71	

TABLE 2
STATUS OF FGD SYSTEMS DURING 1974

COUNTRY	FGD STATION	CURRENT MONTH
DETROIT	1.0. BURNER 31	INSTALLATION OF THE FGD SYSTEM IS ESSENTIALLY COMPLETE. A FAULTY INSTRUMENT PANEL WHICH WAS INCORRECTLY WIRED HAS BEEN RETURNED TO THE MANUFACTURER FOR REPAIR. ITS DELIVERY IS NOT EXPECTED BEFORE NOVEMBER 15TH.
	2.0. 2.0. - DETROIT	IN THE MEANTIME, DETROIT EDISON HAS WATER TESTED SUCH EQUIPMENT AS PUMPS, TANKS AND PIPING. ALSO THE FGD PANS HAVE BEEN TESTED ON AIR AND THEIR BALANCE CHECKED. STARTUP IS EXPECTED BEFORE THE END OF DECEMBER.
	3.0. 3.0. - DETROIT	
	4.0. 4.0. - DETROIT	
DETROIT	1.0. BURNER 32	EQUIPMENT INSTALLED CONSISTS OF FIVE SINGLE STAGE VENTURIS FOR PARTICULATE REMOVAL. ADDITION OF SECOND STAGE VENTURIS OR POSSIBLY JUST USE OF A HIGHLY REACTIVE LINE TO CONTROL SO ₂ EMISSIONS AWAITS RESULTS OF TESTS AT PHILLIPS.
	2.0. 2.0. - DETROIT	
	3.0. 3.0. - DETROIT	
	4.0. 4.0. - DETROIT	
DETROIT	1.0. BURNER 33	FGD SYSTEM CONSISTS OF 1 TWO STAGE AND 3 ONE STAGE VENTURI SCRUBBERS. AT PRESENT, ONLY FGD GAS FROM FIDELMS 2,3 AND 4 (ABOUT 40% OF THE STATION CAPACITY) ARE IN SERVICE BECAUSE FLY ASH FROM INEFFICIENT ESP'S OVERLOADS THE CLARIFIERS. OPERATING HOURS FOR MODULES 1 THROUGH 4 BETWEEN 3/17/74 AND 3/20/74 WERE 1756, 762, 615, 1707 RESPECTIVELY. PLANT TRIES TO RUN MODULES 1 AND 4 CONTINUOUSLY.
	2.0. 2.0. - DETROIT	
	3.0. 3.0. - DETROIT	
	4.0. 4.0. - DETROIT	
DETROIT	1.0. BURNER 34	THE FGD SYSTEM HAS BEEN DOWN SINCE OCTOBER 4TH AND WILL REMAIN DOWN TILL NOVEMBER 4TH. THE SHUT DOWN WAS DUE TO TROUBLE IN THE CLARIFIER'S NAME MECHANISM. THE NAME'S TURNING GEAR DRIVE WAS DAMAGED BY SOLIDS BUILDUP IN THE GEAR BOX. REPAIRS TO THE CLARIFIER AS WELL AS TO OTHER MINOR PROBLEMS TO THE SYSTEM WILL BE MADE DURING THIS SHUTDOWN. THE FGD SYSTEM IS RUNNING AT HALF CAPACITY (SINCE NAME) DUE TO LOW HEAT DEMANDS.
	2.0. 2.0. - DETROIT	
	3.0. 3.0. - DETROIT	
	4.0. 4.0. - DETROIT	
DETROIT	1.0. BURNER 35	FIVE PROPOSALS FOR FGD SYSTEMS HAVE BEEN RECEIVED AND A DECISION WILL BE MADE SOON. LINE AND LIMESTONE SCRUBBERS ARE THE FRONT RUNNERS BUT COAL DESULFURIZATION IS ALSO BEING CONSIDERED.
	2.0. 2.0. - DETROIT	
	3.0. 3.0. - DETROIT	
	4.0. 4.0. - DETROIT	

TABLE 2
STATUS OF FGD SYSTEMS DURING 1/7/74

Company	Power Station	Current Month
I.G. RUPPER 36	ILLINOIS POWER	THE FGD UNIT HAS BEEN DOWN THE PAST 60 DAYS TO PERFORM SEVERAL REPAIRS. THE REPAIR OF THE REFRACTORY BRICKS IN THE HEATER BURNER DOME HAS BEEN COMPLETED. REPAIRS TO REFRACTORY LINED DUCTS IS IN PROGRESS. OTHER MAJOR REPAIRS CONSIST OF REINFORCEMENT OF THE STRUCTURAL STEEL SUPPORT TO THE HIGH TEMPERATURE STAINLESS STEEL DUCTWORK AND REPAIR OF THE ACID COOLER BAFFLES. THE UNIT IS SCHEDULED TO RESTART BY NOVEMBER 22, 1974.
110 1.0 - RETROFIT	COAL 2.9- 3.2 PERCENT SULFUR	
MORGANTHAU STEEL CO. CHM SYSTEMS	CATALYTIC OXIDATION	
STARTUP 11/72		
I.G. RUPPER 37	INDIANAPOLIS POWER AND LIGHT	INDIANAPOLIS POWER AND LIGHT HAS MET WITH SOME VENDORS IN AN EFFORT TO OBTAIN MORE COMPLETE BIDS. THEY ARE NOW AWAITING REPLIES.
110 1.0 - RETROFIT	COAL 3.1-3.5 PERCENT SULFUR	
STARTUP 4/77		
I.G. RUPPER 38	WABASH CITY POWER & LIGHT	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TO OCTOBER 1ST. DURING THIS PERIOD, THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, BECAUSE OF THE MANPOWER SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT RESTARTED OCTOBER 1ST AND HAS BEEN RUNNING INTERMITTENTLY DURING THIS MONTH WITH AN AVAILABILITY OF ABOUT 40 PERCENT.
110 1.0 - RETROFIT	COAL 1.0- 4.0 PERCENT SULFUR	
STARTUP 11/72		
I.G. RUPPER 39	WABASH CITY POWER & LIGHT	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JUL 8TH TILL OCT 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, DUE TO THE MANPOWER SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT HAS NOT RESTARTED BECAUSE THE COOLER IS DOWN TO OVERHAUL THE TUBING ON THE UNIT'S ELECTRIC GENERATOR.
110 1.0 - RETROFIT	COAL 1.0- 4.0 PERCENT SULFUR	
STARTUP 8/72		
I.G. RUPPER 40	WABASH CITY POWER & LIGHT	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TILL OCT 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. THE FGD SYSTEM OPERATED WITH ABOUT 80 PERCENT AVAILABILITY DURING OCTOBER. THERE ARE NO MAJOR OUTSTANDING PROBLEMS. HOWEVER, PREVENTIVE MAINTENANCE PRACTICE REQUIRES THAT ONE MODULE BE TAKEN DOWN ONCE A WEEK FOR 8 HOURS OF CLEANING AND WASHING OF SOLIDS BUILDUP.
110 1.0 - RETROFIT	COAL 3.2 PERCENT SULFUR	
STARTUP 8/73		

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
POWER STATION 41		
KANSAS POWER & LIGHT	JEFFERY 10.1	TENTATIVE AGREEMENT HAS BEEN REACHED WITH COMBUSTION ENGINEERING TO DO PRELIMINARY FGD SYSTEM DESIGN FOR THIS UNIT AS PART OF THEIR CONTINUED WORK ON THE FGD SYSTEM AT THE LAWRENCE PLANT.
700 % - NEW	COAL	
POWER STATION 42		
KANSAS POWER & LIGHT	JEFFERY 10.2	TENTATIVE AGREEMENT HAS BEEN REACHED WITH COMBUSTION ENGINEERING TO DO PRELIMINARY FGD SYSTEM DESIGN FOR THIS UNIT AS PART OF THEIR CONTINUED WORK ON THE FGD SYSTEM AT THE LAWRENCE PLANT.
700 % - 12%	COAL	
POWER STATION 43		
KANSAS POWER & LIGHT	JEFFERY 10.3	BOTH FGD MODULES ARE IN OPERATION WITH ESSENTIALLY 100 PERCENT AVAILABILITY. MAINTENANCE WASH-UP OF DEFISTERS IS PERFORMED ABOUT ONCE EVERY TWO WEEKS. MAINTENANCE WASH-UP OF REHEATER COIL IS NOW DUE AFTER MORE THAN 12 MONTHS OF OPERATION.
120 % - RETROFIT	COAL 3.5 PERCENT SULFUR	
POWER STATION 44		
KANSAS POWER & LIGHT	JEFFERY 10.4	ROILER 5 HAS BEEN OPERATING ON COAL SINCE OCTOBER 12TH. POOR GAS DISTRIBUTION THROUGH EACH ONE OF THE 8 MODULES AS WELL AS AMONG THE MODULES IS STILL A PROBLEM. COMBUSTION ENGINEERING HAS COMPLETED GAS DISTRIBUTION TESTS AND IS CURRENTLY MODIFYING THE GAS FLOW PATTERN THROUGH THE MODULES. NO RECORD IS KEPT OF THE FGD SYSTEM'S OPERATING HOURS OR SYSTEM AVAILABILITY.
400 % - RETROFIT	COAL 3.5 PERCENT SULFUR	
POWER STATION 45		
KANSAS POWER & LIGHT	JEFFERY 10.5	THE FGD SYSTEM WILL CONSIST OF ONE MODULE TO HANDLE THE COMBINED FLUE GASES FROM THE THREE BOILERS. MINOR DELAYS HAVE EXPERIENCED IN THE DELIVERY OF STRUCTURAL STEEL. CONSTRUCTION STARTED ABOUT SEPTEMBER 1ST AND COMPLETION IS DUE BY MARCH 1975.
400 % - RETROFIT	COAL 3.3 PERCENT SULFUR	
POWER STATION 46		
KANSAS POWER & LIGHT	JEFFERY 10.6	
400 % - RETROFIT	COAL 3.3 PERCENT SULFUR	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/90 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	120 MW - RETROFIT	
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/90 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	107 MW - RETROFIT	
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/90 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	137 MW - RETROFIT	
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/90 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	176 MW - RETROFIT	
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/90 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
	176 MW - RETROFIT	
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	
LOUISVILLE GAS & ELECTRIC	LOUISVILLE GAS & ELECTRIC	A PERMIT FOR CONSTRUCTION OF THE FGD SYSTEM WAS DISAPPROVED BY THE KENTUCKY PUBLIC SERVICE COMMISSION. HOWEVER, THE ISSUE HAS NOT BEEN FINALLY RESOLVED.
	123 MW - RETROFIT	
	COAL 3.5-4.0 PERCENT SULFUR	
	NOT SELECTED	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/79

COMPANY		CURRENT MONTH	
POWER STATION			
10. POWER 51		A PERMIT FOR CONSTRUCTION OF THE FGD SYSTEM WAS DISAPPROVED BY THE KENTUCKY PUBLIC SERVICE COMMISSION. HOWEVER, THE ISSUE HAS NOT BEEN FINALLY RESOLVED.	
LOUISVILLE GAS & ELECTRIC			
Coke No. 6			
277 PA - RETROFIT			
COAL 3.5- 4.0 PERCENT SULFUR			
NOT SELECTED			
LINE SCRUBBING			
STARTUP 6/77			
10. POWER 52		A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/79 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.	
LOUISVILLE GAS & ELECTRIC			
MILL CREEK NO. 1			
330 PA - RETROFIT			
COAL 3.5- 4.0 PERCENT SULFUR			
NOT SELECTED			
LINE SCRUBBING			
STARTUP 6/79			
10. POWER 53		A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/79 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.	
LOUISVILLE GAS & ELECTRIC			
MILL CREEK NO. 2			
330 PA - RETROFIT			
COAL 3.5- 4.0 PERCENT SULFUR			
NOT SELECTED			
LINE SCRUBBING			
STARTUP 6/79			
10. POWER 54		CONTRACT WAS AWARDED TO AMERICAN AIR FILTER. SYSTEM WILL OPERATE CLOSED LOOP ON WATER AND SLUDGE WILL BE STABILIZED.	
LOUISVILLE GAS & ELECTRIC			
MILL CREEK NO. 3			
425 PA - RETROFIT			
COAL 3.5- 4.0 PERCENT SULFUR			
AMERICAN AIR FILTER			
LINE SCRUBBING			
STARTUP 6/77			
10. POWER 55		A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/79 AS THE STARTUP DATE FOR AN FGD SYSTEM.	
LOUISVILLE GAS & ELECTRIC			
MILL CREEK NO. 4			
425 PA - RETROFIT			
COAL 3.5- 4.0 PERCENT SULFUR			
NOT SELECTED			
LINE SCRUBBING			
STARTUP 6/79			

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY

POWER STATION

CURRENT MONTH

1.D. NUMBER 55
LOUISVILLE GAS & ELECTRIC
PADUYS RIVER
COAL 3.0-4.0 PERCENT SULFUR
COMBUSTION ENGINEERING
LIVE SCRAPING
STARTUP 4/73

1.D. NUMBER 57
LOUISVILLE GAS & ELECTRIC
COAL 3.0-4.0 PERCENT SULFUR
COMBUSTION ENGINEERING
LIVE SCRAPING
STARTUP 4/73

1.D. NUMBER 58
LOUISVILLE GAS & ELECTRIC
COAL 3.0-4.0 PERCENT SULFUR
COMBUSTION ENGINEERING
LIVE SCRAPING
STARTUP 4/73

1.D. NUMBER 59
LOUISVILLE GAS & ELECTRIC
COAL 3.0-4.0 PERCENT SULFUR
COMBUSTION ENGINEERING
LIVE SCRAPING
STARTUP 4/73

UTILITY IS WAITING FOR APPROVAL OF A CERTIFICATE TO CONSTRUCT BOILERS NO.3 AND NO.4 BEFORE PROCEEDING WITH PLANS TO INSTALL FGD SYSTEMS.

UTILITY IS WAITING FOR APPROVAL OF A CERTIFICATE TO CONSTRUCT BOILERS NO.3 AND NO.4 BEFORE PROCEEDING WITH PLANS TO INSTALL FGD SYSTEMS.

TABLE 2
STATUS OF FGD SYSTEMS DURING 1976

COMPANY	POWER STATION	CURRENT ADMIN
NEVADA POWER REID GARDNER NO 1	125 MW - PETROBRIL COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 12/75	FGD SYSTEM PERFORMANCE IS SATISFACTORILY WITH NO OUTSTANDING MECHANICAL OR PROCESS PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADEQUATE SUPPLY OF TITANIA (SODIUM CARBONATE UNIT) WHICH RESULTED IN FREQUENT INTERRUPTIONS IN THE OPERATION OF THE FGD SYSTEM. NEVADA POWER IS SUEING THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SODIUM CARBONATE UNIT PURCHASED FROM AN ALTERNATE SUPPLIER.
NEVADA POWER REID GARDNER NO 2	125 MW - PETROBRIL COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 12/75	FGD SYSTEM PERFORMANCE IS SATISFACTORILY WITH NO OUTSTANDING MECHANICAL OR PROCESS PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADEQUATE SUPPLY OF TITANIA (SODIUM CARBONATE UNIT) WHICH RESULTED IN FREQUENT INTERRUPTIONS IN THE OPERATION OF THE FGD SYSTEM. NEVADA POWER IS SUEING THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SODIUM CARBONATE UNIT PURCHASED FROM AN ALTERNATE SUPPLIER.
NEVADA POWER REID GARDNER NO 3	125 MW - PETROBRIL COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 6/75	FGD SYSTEM IS UNDER CONSTRUCTION.
NEVADA POWER REID GARDNER NO 4	125 MW - PETROBRIL COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 6/75	NEVADA POWER COMPANY HAS SIGNED A LETTER OF INTENT WITH COMBUSTION EQUIPMENT ASSOCIATES FOR THE CONSTRUCTION OF AN FGD SYSTEM ON REID GARDNER NO. 4. THIS UNIT IS SCHEDULED FOR COMPLETION IN 1978.
NEVADA POWER REID GARDNER NO 5	125 MW - PETROBRIL COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 6/75	UNIVERSAL OIL PRODUCTS IS CONDUCTING PRELIMINARY DESIGN WORK ON A DRY FGD UNIT WITH SULFUR RECOVERY. DECISION TO BE MADE AROUND NOV 1 WHETHER TO SIGN AN ENGINEERING CONTRACT.
NEVADA POWER REID GARDNER NO 6	125 MW - PETROBRIL COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 07/0	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY	POWER STATION	CURRENT MONTH
NORTHERN INDIANA PUB SERVICE	1.0. TURNER 66	CONSIDERING A LINE ON LIMESTONE SCRUBBING UNIT. ALSO WAITING FOR PERFORMANCE OF WELLMAN LONG UNIT
	EMILY CO. 7	UNDER CONSTRUCTION AT THEIR D.H. MITCHELL NO. 11 UNIT
	190 % - RETROFIT	
	COAL 3 PERCENT SULFUR	
NORTHERN INDIANA PUB SERVICE	NOT SELECTED	
	NOT SELECTED	
	STARTUP 0/0	
	400 % - RETROFIT	
NORTHERN INDIANA PUB SERVICE	COAL 3 PERCENT SULFUR	
	NOT SELECTED	
	NOT SELECTED	
	STARTUP 6/0	
NORTHERN INDIANA PUB SERVICE	1.0. TURNER 67	CONSIDERING A LINE ON LIMESTONE SCRUBBING UNIT. ALSO WAITING FOR PERFORMANCE OF WELLMAN LONG UNIT
	DAILY NO. 4	UNDER CONSTRUCTION AT THEIR D.H. MITCHELL NO. 11 UNIT
	400 % - RETROFIT	
	COAL 3 PERCENT SULFUR	
NORTHERN INDIANA PUB SERVICE	NOT SELECTED	
	NOT SELECTED	
	STARTUP 6/0	
	400 % - RETROFIT	
NORTHERN INDIANA PUB SERVICE	COAL 3.2- 3.5 PERCENT SULFUR	
	NOT SELECTED	
	NOT SELECTED	
	STARTUP 6/75	
NORTHERN INDIANA PUB SERVICE	1.0. TURNER 69	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. PURCHASE ORDERS FOR SCRUBBERS WERE PLACED IN EARLY 1971. RECENT SCRUBBER DESIGN CHANGES HAVE CAUSED EQUIPMENT DELIVERY DELAYS BUT THE OVERALL SCHEDULE HAS NOT BEEN UPSET. PROBLEMS HAVE ARISEN IN OBTAINING SOME ELECTRICAL CONTROLS.
	EMILY CO. 1	
	600 % - RETROFIT	
	COAL 1.0 PERCENT SULFUR	
NORTHERN INDIANA PUB SERVICE	COMPOSITION ENGINEERING	
	LIMESTONE SCRUBBING	
	STARTUP 5/75	
	600 % - RETROFIT	
NORTHERN INDIANA PUB SERVICE	1.0. TURNER 70	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. PURCHASE ORDERS FOR SCRUBBERS WERE PLACED IN EARLY 1971. RECENT SCRUBBER DESIGN CHANGES HAVE CAUSED EQUIPMENT DELIVERY DELAYS BUT THE OVERALL SCHEDULE HAS NOT BEEN UPSET. PROBLEMS HAVE ARISEN IN OBTAINING SOME ELECTRICAL CONTROLS.
	EMILY CO. 2	
	600 % - RETROFIT	
	COAL 1.0 PERCENT SULFUR	
NORTHERN INDIANA PUB SERVICE	COMPOSITION ENGINEERING	
	LIMESTONE SCRUBBING	
	STARTUP 5/77	
	600 % - RETROFIT	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

Company		CURRENT MONTH	
FOUR STATIONS		FGD SYSTEM UNDER CONSTRUCTION, BUT ENCOUNTERING SIX MONTH DELAY IN STRUCTURAL STEEL AND VESSEL DELIVERY.	
I.D. NUMBER	71		
PENNSYLVANIA POWER CO.			
BOILER #1 - 72% #2 - 72%			
COAL 4.3 PERCENT SULFUR			
CHEMICO			
LINE SCRUBBING			
STARTUP 6/75			
I.D. NUMBER	72	FGD SYSTEM UNDER CONSTRUCTION, BUT ENCOUNTERING SIX MONTH DELAY IN STRUCTURAL STEEL AND VESSEL DELIVERY.	
PENNSYLVANIA POWER CO.			
BOILER #1 - 72% #2 - 72%			
COAL 4.3 PERCENT SULFUR			
CHEMICO			
LINE SCRUBBING			
STARTUP 6/75			
I.D. NUMBER	73		
PENNSYLVANIA POWER CO.			
BOILER #1 - 72% #2 - 72%			
COAL 4.3 PERCENT SULFUR			
CHEMICO			
LINE SCRUBBING			
STARTUP 12/74			
I.D. NUMBER	74	THE SYSTEM CONSISTS OF 3 PARALLEL VENTURI SCRUBBERS (ONLY ONE IS FOLLOWED BY A 2-STAGE VENTURI SCRUBBER FOR SO ₂ REMOVAL). THE IN OF MODULES TO THE BOILER HAS STARTED AND IS DUE FOR COMPLETION BY DECEMBER 1ST. THE SYSTEM WILL BE TESTED WITH WATER FOR PARTICULATE REMOVAL AND DEBURRING CURING DECEMBER AND JANUARY. SO ₂ SCRUBBING WILL NOT BEGIN BEFORE FEBRUARY 1ST.	
PENNSYLVANIA ELECTRIC			
EMPHOSOL 70.1			
COAL 2.0 PERCENT SULFUR			
CHEMICO			
LINE SCRUBBING			
STARTUP 12/74			
I.D. NUMBER	75	THE FGD SYSTEM CONSISTS OF ONE VENTURI STAGE FOR FLY ASH SCRUBBING WITH WATER FOLLOWED BY A SECOND VENTURI STAGE FOR SO ₂ SCRUBBING WITH MGO SLURRY. THE SYSTEM WAS OPERATED FROM OCTOBER 2ND TO THE 14TH. PRESENTLY THE SYSTEM IS DOWN TO REPAIR LEAKS IN THE FIRST VENTURI STAGE AS WELL AS PROBLEMS WITH THE MGO'S BUCKET ELEVATOR. AVAILABILITY SINCE AUGUST 20TH AVERAGED 34 PERCENT.	
PENNSYLVANIA ELECTRIC & POWER			
EMPHOSOL 70.1			
COAL 2.0 PERCENT SULFUR			
CHEMICO			
LINE SCRUBBING			
STARTUP 9/75			

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY		CURRENT MONTH	
POWER STATION			
I.G. NUMBER 76 PUBLIC SERVICE OF MICHIGAN GIBSON, NO. 2	650 MW - RETROFIT COAL 1.5 PERCENT SULFUR NOT SELECTED LIMESTONE SCRUBBER STARTUP 5/75	UNIT NO. 2 SHOULD BE COMPLETED BY SUMMER 1975. UTILITY IS EXPERIENCING SOME DIFFICULTY IN OBTAINING CONSTRUCTION MATERIALS AND CHEMICALS. TOTAL UNIT CAPACITY IS 650 MW.	
I.G. NUMBER 77 PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 1	375 MW - NEW COAL 0.8 PERCENT SULFUR DAVE FORK RIVER WELLMAN LUMP STARTUP 11/77	FGD PROJECT IS PROCEEDING ACCORDING TO PLAN WITH NO MAJOR PROBLEMS. EQUIPMENT IS PRESENTLY BEING PROCURED. CONSTRUCTION TO BEGIN IN AUGUST 1975	
I.G. NUMBER 78 PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 2	390 MW - RETROFIT COAL 0.8 PERCENT SULFUR DAVE FORK RIVER WELLMAN LUMP STARTUP 1/77	FGD PROJECT IS PROCEEDING ACCORDING TO PLAN WITH NO MAJOR PROBLEMS. EQUIPMENT IS PRESENTLY BEING PROCURED. CONSTRUCTION TO BEGIN IN AUGUST 1975	
I.G. NUMBER 79 PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 3	500 MW - NEW COAL 0.8 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 5/79	PLANS FOR THE BOILER ARE IN THE ENGINEERING STAGE. WORK ON FGD SYSTEM IS IN THE EARLY PLANNING STAGE.	
I.G. NUMBER 80 PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 4	500 MW - RETROFIT COAL 0.8 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 5/80	PLANS FOR THE BOILER ARE IN THE ENGINEERING STAGE. WORK ON FGD SYSTEM IS IN THE EARLY PLANNING STAGE.	

TABLE 2
STATUS OF FGD SYSTEMS WORKING 10/74

POWER STATION	CURRENT MONTH	STATUS OF FGD SYSTEMS DURING 10/74
1.0. POWER 81 S. CAROLINA FGD STATION AUTHORITY GLUCONIC ACID 140 25 - FGD COAL 1.0 PERCENT SULFUR NOT SELECTED LIME/LIFE/SLACK SCRUBBING STARTUP 5/77	VENTURON NOT SELECTED. UIDS ARE DUE IN OCTOBER 74. THE STATION CAPACITY IS 280 MW. ALTERNATIVE UIDS HAVE BEEN REGULATED FOR THE TOTAL 280 MW CAPAC- ITY AND A 1.9% SULFUR COAL.	TABLE 2
1.0. POWER 82 SALT RIVER PROJECT NAVAGO 10 1 750 14 - 0.5 COAL 0.5 - 0.8 PERCENT SULFUR NOT SELECTED LIME/LIFE/SLACK SCRUBBING STARTUP 5/76	AWAITING TEST RESULTS ON HORIZONTAL AND VERTICAL SCRUBBING SYSTEMS AT MONAHE. CONTRACT WILL BE AWARDED 6/1/75.	
1.0. POWER 83 SALT RIVER PROJECT NAVAGO 10 2 750 14 - 0.5 COAL 0.5 - 1.0 PERCENT SULFUR NOT SELECTED LIME/LIFE/SLACK SCRUBBING STARTUP 10/76	AWAITING TEST RESULTS ON HORIZONTAL AND VERTICAL SCRUBBING SYSTEMS AT MONAHE. CONTRACT WILL BE AWARDED 6/1/75.	
1.0. POWER 84 SALT RIVER PROJECT NAVAGO 10 3 750 14 - 0.5 COAL 0.5 - 0.8 PERCENT SULFUR NOT SELECTED LIME/LIFE/SLACK SCRUBBING STARTUP 5/77	AWAITING TEST RESULTS ON HORIZONTAL AND VERTICAL SCRUBBING SYSTEMS AT MONAHE. CONTRACT WILL BE AWARDED 6/1/75.	
1.0. POWER 85 SALT RIVER PROJECT NAVAGO 10 4 750 14 - 0.5 COAL 0.5 - 0.8 PERCENT SULFUR NOT SELECTED LIME/LIFE/SLACK SCRUBBING STARTUP 5/77	THE TOTAL CAPACITY OF MILLER IS 800 MW. EXISTING EXPERIMENTAL FGD MODULE WILL TRENT ONLY 160 MW. FINAL SELECTION OF THE PROCESS IS PENDING THE OUTCOME OF TEST RUNS NOW IN PROGRESS AT THE MONAHE STATION.	

TABLE 2
STATUS OF FGD SYSTEMS DURING 10/74

COMPANY		CURRENT MONTH	
POWER STATION			
I.O. POWER CO. SOUTHERN CALIFORNIA Edison MOHAVE NO. 1	100 MW - RETROFIT	THE REPAIRS WHICH WERE PLANNED FOR THE FIRE DAMAGED EQUIPMENT ON THE VERTICAL MODULE ARE ESSENTIALLY COMPLETE. THE SYSTEM HAS BEEN CHECKED OUT USING AMBIENT AIR AND WATER. THE FINAL PHASE OF STARTUP USING FLUE GAS AND LIMESTONE SLURRY IS NOW IN PROGRESS.	
	COAL 0.5- 0.8 PERCENT SULFUR		
	UNIVERSAL OIL PRODUCTS		
	LIMESTONE SCRUBBER STARTUP 12/74		
I.O. POWER CO. SOUTHERN CALIFORNIA Edison MOHAVE NO. 2	100 MW - RETROFIT	THE TEST RUNS UNDER THE NAVAJO PLANT OPERATING CONDITIONS HAVE BEEN COMPLETED. THIS EXPERIMENTAL HORIZONTAL MODULE HAS BEEN TEST RUN ON LIMESTONE FOR ABOUT TWO WEEKS IN OCTOBER. THE UNIT IS NOW BACK ON LINE. THE AVERAGE AVAILABILITY OF THIS FGD SYSTEM SINCE ITS INITIAL START UP IN JANUARY 1974 IS ABOUT 80 PERCENT.	
	COAL 0.5- 0.6 PERCENT SULFUR		
	SLURRY SYSTEM BACK		
	LINE SCRAMBLING STARTUP 11/72		
I.O. POWER CO. SOUTHERN CALIFORNIA Edison MOHAVE NO. 3	100 MW - RETROFIT	THE TOTAL CAPACITY OF BOILER IS 800 MW. EXISTING EXPERIMENTAL FGD MODULE WILL TREAT ONLY 160 MW. FINAL SELECTION OF THE PROCESS IS PENDING THE OUTCOME OF TEST RUNS NOW IN PROGRESS AT THE MOHAVE STATION.	
	COAL 0.5- 0.6 PERCENT SULFUR		
	NOT SELECTED		
	LINE SCRAMBLING STARTUP 6/77		
I.O. POWER CO. SOUTHERN MISSISSIPPI P&H COOP BATTELSBURG NO. 1	223 MW - NEW	FGD SYSTEM IS IN EARLY PLANNING STAGE. UTILITY COMPANY'S CONSULTANT, BURNS & MCDONNELL, IS PRESENTLY PREPARING SPECIFICATIONS FOR BIDS DOCUMENTS.	
	COAL		
	NOT SELECTED		
	LIMESTONE SCRUBBER STARTUP 6/79		
I.O. POWER CO. SOUTHERN MISSISSIPPI P&H COOP BATTELSBURG NO. 2	223 MW - NEW	FGD SYSTEM IS IN EARLY PLANNING STAGE. UTILITY COMPANY'S CONSULTANT, BURNS & MCDONNELL, IS PRESENTLY PREPARING SPECIFICATIONS FOR BIDS DOCUMENTS.	
	COAL		
	NOT SELECTED		
	LIMESTONE SCRUBBER STARTUP 6/79		

TABLE 2
STATUS OF FGD SYSTEMS UNDER 15/74

COUNTRY	POWER STATION	CURRENT MONTH
UNITED STATES	1.0. WADSWORTH 91 SOUTHERN PUBLIC SERVICE HARRINGTON RD 1 345 PA - 540 COAL 0.5 PERCENT SULFUR CONNECTIONS ESTABLISHED LIVESTOCK SCOURING STARTUP 8/78	PLANT IS UNDER CONSTRUCTION BUT EXPERIENCING DELAYS IN DELIVERY OF STRUCTURAL STEEL AND ALLOY PIPES AND VALVES. EXPECT TO HAVE STRUCTURAL STEEL FOR SCOURERS ERECTED BY 3RD QUARTER 1975.
	1.0. WADSWORTH 92 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	THE BOILER IS UNDER CONSTRUCTION. INSTALLATION OF FGD SYSTEM WILL START IN NOVEMBER 74. PRESENTLY, THE FGD PROJECT IS IN ENGINEERING AND EQUIPMENT PROCUREMENT STAGE.
	COAL LIVESTOCK SCOURING STARTUP 8/78	
	1.0. WADSWORTH 93 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	THIS IS AN EXPERIMENTAL SYSTEM FUNDED BY USEPA. THERE ARE THREE TYPES OF SCOURERS (10% EQUIVALENT EACH) BEING EVALUATED UNDER A VARIETY OF OPERATIONAL MODELS.
UNITED STATES	1.0. WADSWORTH 94 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	ALL SCOUR PROCESS EQUIPMENT HAS BEEN PURCHASED. IN-LINE HEATER SCRAPPED AND DIRECT HEAT SYSTEM IS IN DESIGN STAGE.
	COAL LIVESTOCK SCOURING STARTUP 1/77	
	1.0. WADSWORTH 95 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	
	COAL LIVESTOCK SCOURING STARTUP 1/77	
UNITED STATES	1.0. WADSWORTH 96 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	
	COAL LIVESTOCK SCOURING STARTUP 1/77	
	1.0. WADSWORTH 97 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	
	COAL LIVESTOCK SCOURING STARTUP 1/77	
UNITED STATES	1.0. WADSWORTH 98 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	
	COAL LIVESTOCK SCOURING STARTUP 1/77	
	1.0. WADSWORTH 99 SOUTHERN PUBLIC SERVICE SOUTHWEST RD 1 200 PA - 540	
	COAL LIVESTOCK SCOURING STARTUP 1/77	

TABLE 2
STATUS OF FGD SYSTEMS DURING 1977

CURRENT MONTH

PG&E STATION

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

1.0. 10.00 96

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	17
UTILITY NAME	BOSTON EDISON
UNIT NAME	WYSLIC NO 6
UNIT LOCATION	LYNNETT MASSACHUSETTS
UNIT RATING	150 MW
FUEL CHARACTERISTICS	OIL 2.5 PERCENT SULFUR
FGD VENDOR	CHEMICO
PROCESS	WAGNER/STUR OXIDE SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	9/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	50 PERCENT
SO ₂	90 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	
UNIT COST	\$25/MW CAPITAL+3.9 MILLS/MW OP
OPERATIONAL EXPERIENCE	FGD SYSTEM IS DOWN WITH NO IMMEDIATE PLANS FOR RESTARTING BECAUSE EPA FUNDING HAS EXPIRED AND THE CALCINING PLANT IS NOW BEING USED BY ANOTHER COMPANY. THE UTILITY IS EVALUATING THE DATA COLLECTED DURING THE DEMON- STRATION PERIOD TO DETERMINE THE COURSE OF FUTURE ACTION.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	22	CITY OF KEY WEST
UTILITY NAME	KEY WEST POWER PLANT	
UNIT NAME	KEY WEST FLORIDA	
UNIT LOCATION	37 MW	
FUEL CHARACTERISTICS	OIL 2.4 PERCENT SULFUR	
FGD METHOD	ZINC AIR SYSTEMS	
PROCESS	LIMESTONE SCRUBBING	
NEW OR RE-PORT	NEW	
START UP DATE	10/72	
FGD STATUS	OPERATIONAL	
88 EFFICIENCY, PERCENTAGES	90 PERCENT	
SUE	85 PERCENT ESTIMATED	
WATER MAKE UP	5000 LBS RATE ESTIMATED 100 GPM	
SULFUR DISPOSAL	CELLULOSE PAPER	
UNIT COST		
OPERATIONAL EXPENSE	PLANT HAS BEEN SHUT DOWN SINCE SEPT 1974 BECAUSE OF A BOILER I D FAN NO- TICE FAILURE. THEY HOPE TO RE OPERATIONAL THE WEEK OF OCTOBER 21ST. MECH- NICAL FAILURE IN THE FGD SYSTEM WERE FIXED. FGD OPERATED ABOUT A WEEK AND A HALF BEFORE THE PRESENT BOILER SHUTDOWN. DURING SHUTDOWN THE EX- HAUSTION JOINTS OF THE DUCTWORK TO THE SCRUBBER WERE REPLACED FOR THE 2ND TIME. SOME PLUGGING REMOVED AT SHUTDOWN.	

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	29
UTILITY NAME	CONSUMERS HEALTH EDISON
UNIT NAME	WILL COUNTY NO 1
UNIT LOCATION	ROMEOVILLE ILLINOIS
UNIT RATING	167 MW
FUEL CHARACTERISTICS	COAL 0.6-3.0 PERCENT SULFUR
FGD VENDOR	BABCOCK & WILCOX
PROCESS	LIMESTONE SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	2/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	98 PERCENT
SO ₂	75-80 PERCENT-AVERAGE
WATER MAKE UP	106 GAL/LB MOLE SO ₂ REMOVED
SUDGE DISPOSAL	STABILIZED SUDGE DISPOSED IN CLAY LINED POND
UNIT COST	\$96/KW CAPITAL+4.5 MILS/KWH OP
OPERATIONAL EXPERIENCE	MODULE A WAS DOWN 3 TIMES LAST MONTH* DUE TO CLEAR DEPOSITS FROM THE VENTURI THROAT AND TWICE DUE TO BOILER RELATED TROUBLES. DEMISTER CONDITION DETRIMENTED SLIGHTLY AND WILL HAVE TO BE HAND WASHED DURING SCHEDULED BOILER OUTAGE. OPERATION MOST OF THE MONTH WAS WITH A MEDIUM SULFUR COAL BLEND. AVAILABILITY WAS 65 PERCENT. MODULE B IS STILL DOWN.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	50
UTILITY TYPE	KIRKLAND POWER COOP
UNIT NAME	ALMA STATION
UNIT LOCATION	ALMA WISCONSIN
UNIT RATING	60 MW
FUEL CHARACTERISTICS	COAL 3.0- 3.5 PERCENT SULFUR
FGD VENDOR	FOSTER WHEELER
PROCESS	LINE INJECTION
NEW OR REBUILT	REBUILT
START UP DATE	6/71
FGD STATUS	OPERATIONAL
EFFICIENCY, PERCENT	
SUL	25 PERCENT
WATER GASE UP	
SOLIDS DISPOSAL	
UNIT COST	
OPERATIONAL EXPERIENCE	EXPERIENTIAL W/ LINE FURNACE INJECTION SYSTEM WITH NO TAIL END SCRUBBING IS USED. OUTSTANDING PROBLEMS ARE FURNACE TUBE PLUGGAGE WHICH NECESSITATES MAJOR CLEANUP AND BLOW OUT LINE GRINDING AND INJECTION EQUIPMENT. REMOVAL EFFICIENCY WAS FROM 25 TO 40 PERCENT. THE FGD SYSTEM DID NOT RUN MORE THAN TWO DAYS AT A TIME. AVAILABILITY SINCE STARTUP NEVER EXCEEDED 25 PERCENT.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	33
UTILITY NAME	DUQUESNE LIGHT
UNIT NAME	PHILLIPS
UNIT LOCATION	SOUTH HEIGHT PENNSYLVANIA
UNIT RATING	410 MW
FUEL CHARACTERISTICS	COAL 1.0- 2.8 PERCENT SULFUR
FGD MEDIA	CHEMICAL
PROCESS	LINE SCRUBBING
NEW OR REIPOFIT	RETROFIT
START UP DATE	7/73
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	
WATER MAKE UP	
SLUDGE DISPOSAL	STABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$70-75/KW CAP. \$2.1 MILS/KWH OP
OPERATIONAL EXPERIENCE	FGD SYSTEM CONSISTS OF 1 TWO STAGE AND 3 ONE STAGE VENTURI SCRUBBERS. AT PRESENT, ONLY FLUE GAS FROM BOILERS 2,3 AND 4 (ABOUT 40% OF THE STATION CAPACITY) ARE IN SERVICE BECAUSE FLY ASH FROM INEFFICIENT ESP'S OVERLOADS THE CLARIFIERS. OPERATING HOURS FOR MODULES 1 THROUGH 4 BETWEEN 3/17/74 AND 8/30/74 WERE 1756,762,815,1707 RESPECTIVELY. PLANT TRIES TO RUN MODULES 1 AND 4 CONTINUOUSLY.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 1974

IDENTIFICATION NO.	34
UTILITY NAME	GENERAL MOTORS
UNIT NAME	CLEVELAND PARK 1 2 3 & 4
UNIT LOCATION	PARTIA OHIO
UNIT RATING	32 MW
FUEL CHARACTERISTICS	COAL 2.5 PERCENT SULFUR
FGD VEHICLE	ROCK
PROCESS	DOUBLE ALKALI
NEW OR RETROFIT	RETROFIT
START UP DATE	3/74
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO2	
WATER MAKE UP	
SLUDGE DISPOSAL	
UNIT COST	
OPERATIONAL EXPERIENCE	THE FGD SYSTEM HAS BEEN DOWN SINCE OCTOBER 4TH AND WILL REMAIN DOWN TILL NOVEMBER 4TH. THE SHUT DOWN WAS DUE TO TROUBLE IN THE CLARIFIER'S RAKE MECHANISM. THE RAKE'S TURBINE GEAR DRIVE WAS DAMAGED BY SOLIDS BUILDUP IN THE GEAR BOX. REPAIRS TO THE CLARIFIER AS WELL AS TO OTHER MINOR PROBLEMS TO THE SYSTEM WILL BE MADE DURING THIS SHUTDOWN. THE FGD SYSTEM IS RUNNING AT HALF CAPACITY (SINCE MARCH) DUE TO LOW HEAT DEMANDS.

Table 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	36
UTILITY NAME	ILLINOIS POWER
UNIT NAME	WOOL RIVER NO 4
UNIT LOCATION	EAST ALTON ILLINOIS
UNIT RATING	110 Mw
FUEL CHARACTERISTICS	COAL 2.9- 3.2 PERCENT SULFUR
FGD VENDOR	MONSANTO ENVIRO CHEM SYSTEMS
PROCESS	CATALYTIC OXIDATION
ASP OR RELIEF IT	RELIEF IT
START UP DATE	10/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	85 PERCENT
WATER REUSE UP	
SOLID DISPOSAL	
UNIT COST	\$73/KK CAPTured 44 MILLS/KWH UP
OPERATIONAL EXPERIENCE	THE FGD UNIT HAS BEEN DOWN THE PAST 60 DAYS TO PERFORM SEVERAL REPAIRS. THE REPAIR OF THE REFRACTORY BRICKS IN THE REHEATER DUNNEN DOME HAS BEEN COMPLETED. REPAIRS TO REFRACTORY LINED DUCTS IS IN PROGRESS. OTHER MAJOR REPAIRS CONSIST OF REINFORCEMENT OF THE STRUCTURAL STEEL SUPPORT TO THE HIGH TEMPERATURE STAINLESS STEEL DUCTWORK AND REPAIR OF THE ACID COOLER BAFFLES. THE UNIT IS SCHEDULED TO RESTART BY NOVEMBER 22, 1974.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10774

IDENTIFICATION NO.	38
UTILITY NAME	KANSAS CITY POWER & LIGHT
UNIT NAME	HARTHORN NO 3
UNIT LOCATION	KANSAS CITY MISSOURI
UNIT RATING	140 MW
FUEL CHARACTERISTICS	COAL 0.6- 3.0 PERCENT SULFUR
FGD METHOD	COMBUSTION ENGINEERING
PROCESS	LIMESTONE INJECTION WET SCRUB
NEW OR RETROFIT	RETROFIT
START UP DATE	11/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	99 PERCENT
SO ₂	70 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$19/KW CAPITAL+2.5 MILS/KWH OP
OPERATIONAL EXPERIENCE	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TO OCTOBER 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, BECAUSE OF THE MAINTENANCE SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT RESTARTED OCTOBER 1ST AND HAS BEEN RUNNING INTERMITTENTLY DURING THIS MONTH WITH AN AVAILABILITY OF ABOUT 40 PERCENT.

Table 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	59
UTILITY NAME	KANSAS CITY POWER & LIGHT
UNIT NAME	HAWTHORN NO 4
UNIT LOCATION	KANSAS CITY MISSOURI
UNIT RATING	100 MW
FUEL CHARACTERISTICS	COAL 0.6- 3.0 PERCENT SULFUR
FGD DESIGN	COMBUSTION ENGINEERING
PROCESS	11% STRENGTH AMOUNT SORBED
DESIGN EFFECT	95 PERCENT
START UP DATE	8/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	99 PERCENT
SO ₂	70 PERCENT
WATER MAKE UP	
SOLIDS DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$19/MW CAPITAL • 2.2 MILS/MWH OP
OPERATIONAL EXPERIENCE	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JUL 8TH TILL OCT 1ST. DURING THIS PERIOD THE BOILERS WERE OPERATED BY THE COMPANY'S SUPERVISORY STAFF. HOWEVER, DUE TO THE MANPOWER SHORTAGE, THE FGD UNITS WERE DOWN DURING THE STRIKE. THE FGD UNIT HAS NOT RESTARTED BECAUSE THE BOILER IS DOWN TO OVERHAUL THE TURBINE ON THE UNIT'S ELECTRIC GENERATOR.

Table 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	40
UTILITY NAME	KANSAS CITY POWER & LIGHT
UNIT NAME	LA CYGNE NO 1
UNIT LOCATION	LA CYGNE KANSAS
UNIT RATING	820 MW
FUEL CHARACTERISTICS	COAL 5.2 PERCENT SULFUR
FGD METHOD	HABCOCK & WILCOX
PROCESS	LIMESTONE SCRUBBING
NEW OR RETROFIT	NEW
START UP DATE	6/73
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	98 PERCENT
SO2	72 PERCENT
WATER MAKE UP	
SLUDGE DISPOSAL	UNSTABILIZED SLUDGE DISPOSED IN UNLINED POND
UNIT COST	\$432/M CAPITAL
OPERATIONAL EXPERIENCE	KCP&L HAD A LABOR STRIKE AT THE PLANT WHICH LASTED FROM JULY 8TH TILL OCT. 1ST. DURING THIS PERIOD THE BOILER WAS OPERATED BY THE COMPANY'S SUPERVISORY STAFF. THE FGD SYSTEM OPERATED WITH ABOUT 60 PERCENT AVAILABILITY DURING OCTOBER. THERE ARE NO MAJOR OUTSTANDING PROBLEMS. HOWEVER, PREVENTIVE MAINTENANCE PRACTICE REQUIRES THAT ONE MODULE BE TAKEN DOWN ONCE A WEEK FOR 8 HOURS OF CLEANING AND WASHING OF SOLIDS BUILDUP.

TABLE 5
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	43
UTILITY NAME	MAHON POWER & LIGHT
UNIT NAME	LOWMOORE NO. 4
UNIT LOCATION	LOWMOORE WAREHOUSE
UNIT RATING	125 MW
FUEL CHARACTERISTICS	ANAL 3.5 PERCENT SULFUR
FGD METHOD	CUSTOMER FGD, REBIRING
PROCESS	LIMESTONE INJECTION WET SCRUB
DATE OF REPORT	12/74
START-UP DATE	12/74
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	95 PERCENT
SOL	85 PERCENT
WATER WASH UP	
SOLUBLE DISPOSAL	UNSTABILIZED SOLUBLE DISPOSED IN UNLINED POND
UNIT COST	
OPERATIONAL EXPERIENCE	NO. 4 FGD MODULES ARE IN OPERATION WITH ESSENTIALLY 100 PERCENT AVAILABILITY. MAJOR WASH-UP OF DEVICES IS PERFORMED ABOUT ONCE EVERY TWO WEEKS. MAJOR CLEANUP OF REHEATER COIL IS NOW DONE AFTER MORE THAN 12 MONTHS OF OPERATION.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	44
UTILITY NAME	KANSAS POWER & LIGHT
UNIT NAME	LAWRENCE NO 5
UNIT LOCATION	LAWRENCE KANSAS
UNIT RATING	900 MW
FUEL CHARACTERISTICS	COAL 3.5 PERCENT SULFUR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIVESTONE INJECTION WET SCRUB
NEW OR RETROFIT	NEW
START UP DATE	11/71
FGD STATUS	OPERATIONAL
EFFICIENCY, PERCENTAGES	92 PERCENT
SO ₂	65 PERCENT
WATER MAKE UP	
SLOUGH DISPOSAL	DESIGN LIZED SLOUGH DISPOSED IN UNLINED POND
UNIT COST	
OPERATIONAL EXPERIENCE	UNIT 5 HAS BEEN OPERATING ON COAL SINCE OCTOBER 12TH. POOR GAS DISTRIBUTION THROUGH EACH ONE OF THE 4 MODULES AS WELL AS AMONG THE MODULES IS STILL A PROBLEM. COMBUSTION ENGINEERING HAS COMPLETED GAS DISTRIBUTION TESTS AND IS CURRENTLY MODIFYING THE GAS FLOW PATTERN THROUGH THE MODULES. NO MERCURY IS KEPT OF THE FGD SYSTEM'S OPERATING HOURS OR SYSTEM AVAILABILITY.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	56
UTILITY NAME	LOUISVILLE GAS & ELECTRIC
UNIT NAME	PALDYS RUN NO. 6
UNIT LOCATION	LOUISVILLE KENTUCKY
UNIT RATING	65 MW
FUEL CHARACTERISTICS	COAL 3.5-4.0 PERCENT SULFOR
FGD VENDOR	COMBUSTION ENGINEERING
PROCESS	LIME SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	4/73
FGD STATUS	OPERATIONAL
EFFICIENCY: PARTICULATES	
SO ₂	90 PERCENT RANGE
WATER NAME UP	33 GAL/YD FOLE SO ₂ REMOVED
SLODGE DISPOSAL	INSTABILIZED SLODGE DISPOSED IN UNLINED POND
UNIT COST	\$57/KW CAPITAL
OPERATIONAL EXPENDITURE	FGD SYSTEM OPERATING SATISFACTORILY THE PAST 3 YRS ON LIMESTONE (UNIT OFF FOR REPAIRS). THESE BUNS ARE MADE FOR AND FILLED BY COMBUSTION ENGINEER- ING. AVAILABILITY WAS ESSENTIALLY 100 PERCENT WITH NO PROBLEMS EXPERI- ENCED IN ANY PART OF THE SYSTEM. THE LIMESTONE SLODGE IS DEWATERED IN THE EXISTING ROTARY VACUUM FILTERS WITH NO UNLOADING OF FILTER CLOTH.

Table 3
 FUEL OIL SUPPLY FOR OPERATIONAL FOG SYSTEMS 1974

IDENTIFICATION NO.	61
UTILITY NAME	NEVADA POWER
UNIT DATA	RED GARREN NO 1
UNIT LOCATION	VALPA NEVADA
UNIT RATING	125 Mw
FUEL CHARACTERISTICS	COAL 9.5- 1.0 PERCENT SULF
FUEL SUPPLIER	COMBUSTION EQUIP. ASSOCIATES
PROCESS	SOLUBLE CEMENTATE SCHED 160
NEW AND RETROFIT	RETROFIT
START UP DATE	12/73
FOG STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO2	
WATER SEAL OF	
SOLUBLE DISPOSAL	
UNIT COST	
OPERATIONAL EXPERIENCE	FOG SYSTEM PERFORMANCE IS SATISFACTORY WITH NO OUTSTANDING MECHANICAL OR FUELING PROBLEMS TO REPORT. HOWEVER, THE PLANT IS EXPERIENCING AN INADE- QUATE SUPPLY OF FINEST ISOLATION CARBONATE GEL WHICH RESULTS IN FREQUENT INTERRUPTIONS IN THE OPERATION OF THE FOG SYSTEM. NEVADA POWER IS SOLVING THE SUPPLIER FOR BREACH OF CONTRACT AND ALSO IS PRESENTLY TESTING SOLUBLE CARBONATE GELING PURCHASED FROM AN ALTERNATE SUPPLIER.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS - 10/76

IDENTIFICATION NO.	75
UTILITY NAME	POTOMAC ELECTRIC & POWER
UNIT NAME	DICKERSON NO. 3
UNIT LOCATION	DAVENPORT PARKLAND
UNIT RATING	100 MW
FUEL CHARACTERISTICS	COAL 2.0 PERCENT SULFUR
FGD VENDOR	CH2M HILL
PROCESS	WATERGAS SHIFT
NEW OR REFURBISH	REFURBISH
START UP DATE	9/75
FGD STATUS	OPERATIONAL
EFFICIENCY, PERCENTAGES	
SO ₂	
WATER VAPOR UP	
SOLIDS DISPOSAL	
UNIT COST	\$70/HR CAPITAL
OPERATIONAL EXPERIENCE	THE FGD SYSTEM CONSISTS OF ONE VENTURI STAGE FOR FLY ASH SCRUBBING WITH WATER FOLLOWED BY A SECOND VENTURI STAGE FOR SO₂ SCRUBBING WITH FGD REAGENT. THE SYSTEM WAS OPERATED FROM OCTOBER 1975 TO THE PRESENT. BY THE SYSTEM IS DOWN TO REPAIR LEAKS IN THE FIRST VENTURI STAGE AS WELL AS PROBLEMS WITH THE REAGENT TANK ELEVATOR. AVAILABILITY SINCE AUGUST 1976 AVERAGED 94 PERCENT.

Table 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/74

IDENTIFICATION NO.	67
UTILITY NAME	SOUTHERN CALIFORNIA Edison
UNIT NAME	MOHAVE NO. 2
UNIT LOCATION	LAGOONVIEW NEVADA
UNIT RATING	160 MW
FUEL CHARACTERISTICS	COAL 0.5- 0.8 PERCENT SULFUR
FGD VENTILATOR	SO ₂ / STERIL ROGER
PROCESS	LIME SCRUBBING
NEW OR RETROFIT	RETROFIT
START UP DATE	11/73
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	90 PERCENT
ENTER MAKE UP	370 GAL/LE MOLE SO ₂ REMOVED
SLOUDGE DISPOSAL	STABILIZED SLOUDGE DISPOSED IN UNLINED POND
UNIT COST	
OPERATIONAL EXPENSE	THE TEST RUNS UNDER THE MOHAVE PLANT OPERATING CONDITIONS HAVE BEEN COMPLETED. THIS EXPERIMENTAL HORIZONTAL MODULE HAS BEEN TEST RUN ON LIMESTONE FOR ABOUT TWO WEEKS IN OCTOBER. THE UNIT IS NOW BACK ON LIME. THE AVERAGE AVAILABILITY OF THIS FGD SYSTEM SINCE ITS INITIAL START UP IN JANUARY 1974 IS ABOUT 80 PERCENT.

TABLE 3
PERFORMANCE SUMMARY FOR OPERATIONAL FGD SYSTEMS 10/79

IDENTIFICATION NO.	93
UTILITY NAME	TEHESSEL VALLEY AUTHORITY
UNIT NAME	SHAWNEE NO. 10
UNIT LOCATION	FRANCIS KENTUCKY
UNIT RATING	30 MW
FUEL CHARACTERISTICS	COAL
FGD DESIGN	TEHESSEL VALLEY AUTHORITY
PROCESS	WET/LIME/STONE SCRUBBING
RAW GAS BENEFIT	NO BENEFIT
START OF DATE	9/72
FGD STATUS	OPERATIONAL
EFFICIENCY, PARTICULATES	
SO ₂	
WATER WASH UP	
SLOUGH DISPOSAL	
UNIT COST	
OPERATIONAL EXPERIENCE	THIS IS AN EXPERIMENTAL SYSTEM FUNDED BY USEPA. THERE ARE THREE TYPES OF SCRUBBERS (10 MW EQUIVALENT EACH) BEING EVALUATED UNDER A VARIETY OF OPERATIONAL MODELS.

TABLE 4
NUMBER AND TOTAL NO OF FGD SYSTEMS

STATUS	NO OF UNITS	%
OPERATING	19	52.1.
UNDER CONSTRUCTION	10	26.7.
PLANNING CONTRACT AWARDED	9	23.6.
LEAVE OF ABSENCE	2	5.3.
REDESIGN/RECONSTRUCTION	10	26.7.
CONSISTING ONLY FGD SYSTEMS	37	100.
TOTAL	99	267.

TABLE 5
SUMMARY OF FGD SYSTEMS BY COMPANY

UTILITY	TOTAL				OPERATIONAL				CONSTRUCTION				PLANNED				REQUESTING/				CONSIDERING/							
	NO.				PM				MW				CONTRACT				LETTER OF				EVAL. BIDS				FGD SYSTEMS			
	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM	NO.	PM						
ALABAMA ELECTRIC COOP	2	450.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	2	450.						
ARIZONA ELECTRIC POWER	2	405.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
ARIZONA ELECTRIC SERVICE	2	360.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
ARIZONA ELECTRIC	3	1650.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	6	2550.						
ARIZONA ELECTRIC	4	150.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	3	1650.						
CENTRAL ILLINOIS LIGHT CO.	1	100.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
CENTRAL ILLINOIS PUBLIC SERV	1	600.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
CINCINNATI GAS & ELECTRIC	2	1100.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
CITY OF NEW YORK	1	57.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	1	600.						
COLUMBIA GAS ELECTRIC	4	1350.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
COLUMBIA GAS ELECTRIC	2	750.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
COLUMBIA GAS ELECTRIC	1	167.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DELTA ELECTRIC COOP	1	80.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	180.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	2	920.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	32.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	650.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	110.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	515.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	3	1060.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	4	1925.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	64.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	2507.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	4	2120.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	500.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	650.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	3	700.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	2	1360.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	3	2400.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC	1	120.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
DETROIT ELECTRIC & POWER	1	100.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
PUBLIC SERVICE OF NEW YORK	1	650.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
PUBLIC SERVICE OF NEW YORK	4	1715.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	1	140.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	3	2250.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	4	1600.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	2	446.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	1	345.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	1	200.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	2	500.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
SOUTHERN GAS & ELECTRIC AUTHORITY	5	3972.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.	0	0.						
TOTALS	99	37961.	19	3291.	16	6877.	9	3904.	3	530.	13	6887.	37	16472.														

TABLE 6
SUMMARY OF FGD SYSTEMS BY VENDOR

MANUFACTURER/PROCESS	TOTAL		OPERATIONAL		CONSTRUCTION	
	NO.	FM	NO.	FM	NO.	FM
AMERICAN AIR FILTER						
LINE SCRUBBING	1	84.	0	0.	1	64.
TOTAL, AMERICAN AIR FILTER	1	84.	0	0.	1	64.
BARCOCK & WILCOX						
LIMESTONE SCRUBBING	2	907.	2	987.	0	0.
TOTAL, BARCOCK & WILCOX	2	907.	2	987.	0	0.
CORBICO						
LINE SCRUBBING	4	2600.	1	410.	3	2190.
WATERWASH DRY SCRUBBING	2	250.	2	250.	0	0.
TOTAL, CORBICO	6	2850.	3	660.	3	2190.
CONQUESTION ENGINEERING						
LINE SCRUBBING	1	65.	1	65.	0	0.
LIME/ALUMINA DRY SCRUBBING	1	43.	0	0.	1	343.
LIMESTONE INJECTION DRY SCRUB	4	765.	4	765.	0	0.
LIMESTONE SCRUBBING	2	1350.	0	0.	2	1360.
TOTAL, CONQUESTION ENGINEERING	6	2533.	5	830.	3	1703.
CORROSION EQUIP. ASSOCIATES						
LINE SCRUBBING	2	720.	0	0.	2	720.
SECTOR CARTRIDGE SCRUBBING	3	375.	2	250.	1	125.
TOTAL, CORROSION EQUIP. ASSOCIATES	5	1095.	2	250.	3	845.
DAVY POWERGAS						
WELLBURN LOOP	1	115.	0	0.	1	115.
TOTAL, DAVY POWERGAS	1	115.	0	0.	1	115.
FOSTER WHEELER						
LINE INJECTION	1	80.	1	80.	0	0.
TOTAL, FOSTER WHEELER	1	80.	1	80.	0	0.
KROH						
DOUBLE ALKALI	1	52.	1	52.	0	0.
TOTAL, KROH	1	52.	1	52.	0	0.
KONSATO ENVIRO CHEM SYSTEMS						
CATALYTIC OXIDATION	1	110.	1	110.	0	0.
TOTAL, KONSATO ENVIRO CHEM SYS	1	110.	1	110.	0	0.

TABLE 6
SUMMARY OF FFG SYSTEMS BY VENDOR

MANUFACTURER/PROCESS	STATUS			
	TOTAL	OPERATIONAL	CONSTRUCTION	
	NO.	NO.	NO.	NO.
LOT SELECTED				
LIME/LIMESTONE SCRUBBING	1 650.	0 0.	1 650.	
TOTAL LOT SELECTED	1 650.	0 0.	1 650.	
PLANT/PROCESS				
LIMESTONE SCRUBBING	1 160.	0 0.	1 160.	
TOTAL PLANT/PROCESS	1 160.	0 0.	1 160.	
RESEARCH CENTER				
LIMESTONE SCRUBBING	1 115.	1 115.	0 0.	
TOTAL RESEARCH CENTER	1 115.	1 115.	0 0.	
PILOT SCALE/PLANT/PROCESS				
LIMESTONE SCRUBBING	1 100.	0 0.	1 100.	
TOTAL PILOT SCALE/PLANT/PROCESS	1 100.	0 0.	1 100.	
SEMI SCALE/PLANT/PROCESS				
LIME SCRUBBING	1 160.	1 160.	0 0.	
TOTAL SEMI SCALE/PLANT/PROCESS	1 160.	1 160.	0 0.	
TEMPERATURE VALLEY AUTHORITY				
LIME/LIMESTONE SCRUBBING	1 30.	1 30.	0 0.	
LIMESTONE SCRUBBING	1 550.	0 0.	1 550.	
TOTAL TEMPERATURE VALLEY AUTHORITY	2 580.	1 30.	1 550.	
UNITED ENGINEERS				
LIMESTONE/LIME SCRUBBING	1 120.	0 0.	1 120.	
TOTAL UNITED ENGINEERS	1 120.	0 0.	1 120.	
UNIVERSAL OIL PRODUCTS				
LIMESTONE SCRUBBING	2 360.	0 0.	2 360.	
TOTAL UNIVERSAL OIL PRODUCTS	2 360.	0 0.	2 360.	
ZURU AIR SYSTEMS				
LIMESTONE SCRUBBING	1 37.	1 37.	0 0.	
TOTAL ZURU AIR SYSTEMS	1 37.	1 37.	0 0.	
	37 10166.	19 3291.	18 6877.	

TABLE 5
SUMMARY OF OPERATING FGO SYSTEMS BY
PROCESS AND GENERATING UNITS AS OF 10/74

PROCESS/GENERATING UNITS FGO/yr STARTUP EXPERIENCE (NO.)

CATALYTIC OXIDATION

WOOD RIVER NO 4 110 10-72 24
110 24

DOUBLE ALKALI

CHEVROLET PARMA 1 2 3 & 4 32 3-74 7
52 7

LIME INJECTION

ALMA STATION 80 6-71 40
60 40

LIME SCRUBBING

MOHAVE NO 2 160 11-73 11
PADDOYS RUN NO 6 65 4-73 18
PHILLIPS 410 7-73 15
655 44

LIME/LIMESTONE SCRUBBING

SHAWNEE NO. 10 50 4-72 30
50 30

LIMESTONE SCRUBBING

HAWTHORN NO 3 190 11-72 23
HARTHORN NO 4 100 8-72 26
LAWRENCE NO 4 125 12-68 70
LAWRENCE NO 5 400 11-71 35
CHOLLA NO 1 115 10-73 12

TABLE 6
SUMMARY OF OPERATING FOR SYSTEMS BY
PROCESS AND GENERATING UNITS AS OF 10/79

PROCESS/GENERATING UNITS	EGG/OW	STARTUP	EXPERIENCE (NO.)
KEY WEST POWER PLANT	57	10-72	24
LA CYCLE NO 1	600	6-73	16
WILL COUNTY NO 1	167	2-72	32
	-----		-----
	1904		436
REGRESSION DATE SCHEDULING			
DICKINSON NO 3	100	9-73	13
MYSTIC NO 6	150	4-72	30
	-----		-----
	250		43
SOUTH CAROLINA SCHEDULING			
NEID CARTER NO 1	125	12-73	10
NEID CARTER NO 2	125	12-73	10
	-----		-----
	250		20

TABLE 9
SUMMARY OF SLUDGE DISPOSAL PRACTICES FOR OPERATIONAL FGD SYSTEMS AS OF 10/7/74

PROCESS/GENERATIVE UNIT	---SLUDGE-- STABILIZED	---SLUDGE-- UNSTABILIZED	---POND--- LINED	---POND--- UNLINED
LIME SCRUBBER				
ACHAVE NO 2	160			160
PALCOYS NO 6		85		85
PHILLIPS	410			410
TOTAL	570.	85.	0.	655.
LIMESTONE SCRUBBER				
BAKIDON NO 3		140		140
BAKIDON NO 4		100		100
LAWRENCE NO 4		125		125
LAWRENCE NO 5		400		400
TULLA NO 1		115		115
KEY WEST POWER PLANT				37
LA CYGNE NO 1		820	167	820
WILL COUNTY NO 1	167			
TOTAL	167.	1700.	167.	1737.

TABLE 10

SUMMARY OF FGD WATER TREAT PRACTICES

	PARTIAL	
	OPEN	CLOSED
WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

CATALYTIC CALCULATION

WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

LIME INJECTION

WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

LIME SCRUBBING

WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

113

LIME SCRUBBING

WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

LIME/LIMESTONE SCRUBBING

WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

LIMESTONE INJECTION & LIME SCRUBBING

WACO RIVER TO 4	0	110
CHEVROLET PANTA 1, 2 & 4	0	32
ALMA STATION	0	80
REMOVE TO 2	0	160
PACIFIC RIVER & PHILLIPS	0	65
SHAWNEE NO. 10	0	30
HASTINGS NO. 3	0	0
HASTINGS NO. 4	0	0
LAWRENCE NO. 4	0	0
LAWRENCE NO. 5	0	0

LA 44-10

SUMMARY OF 400 MAILING LOOP PRACTICES

PARTIAL
OPEN CLOSED CLOSED

LIFE-STORE SCHEDULING

COCKLE NO 1	115	0	0
FLY WEST GREEN PLANT	57	0	0
LA CYCLE NO 1	0	420	0
WILL COUNTY NO 1	0	167	0
	152	907	0

MADEIRA CYCLE SCHEDULING

DICKINSON NO 2	0	0	100
MYSTIC NO 6	0	0	150
	0	0	250

STATION CEMENTITE SCHEDULING

WILCO GREEN NO 1	125	0	0
WILCO GREEN NO 2	145	0	0
	270	0	0

Table 14
OPERATIONAL FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	PLANT OR BLDG UNIT	SIZE OF FGD UNIT (MW)	PROCESS/REAGENT	FULL CHARACTERISTICS	NO/YR
ARIZONA PUBLIC SERVICE CHOLLA #0 1	R	115	RESEARCH COUNTELL LIMESTONE SCRUBBING	COAL 0.4% PERCENT SULFUR	10/73
BOSTON Edison MYSTIC #0 6	R	150	CHEMICO MAGNESIUM OXIDE SCRUBBING	OIL 2.5 PERCENT SULFUR	4/72
CITY OF NEW JERSEY NEW JERSEY POWER PLANT	R	37	ZACHRY AIR SYSTEMS LIMESTONE SCRUBBING	OIL 2.4 PERCENT SULFUR	10/72
COMPASS BATH FORDSON WILL COUNTY #0 1	R	167	BRIDGES & MILLER LIMESTONE SCRUBBING	COAL 0.6-3.0 PERCENT SULFUR	2/72
DAKOTA POWER COOP ALMA STATION	R	60	ENTER WHEELER LINE INJECTION	COAL 3.0- 3.5 PERCENT SULFUR	6/71
DEQUESTER LIGHT PHILLIPS	R	410	CHEMICO LINE SCRUBBING	COAL 1.0- 2.0 PERCENT SULFUR	7/73
GENERAL ATOMIC CHRYSLER FERRA 1, 2, 3 & 4	R	52	MOORE DOUBLE ALKALI	COAL 2.5 PERCENT SULFUR	3/74
ILLINOIS POWER SCUD MINE #0 4	R	110	POULSATO RAVING CHEM SYSTEMS CATALYTIC OXIDATION	COAL 2.9- 3.2 PERCENT SULFUR	10/72
KANSAS CITY POWER & LIGHT LA CYCLO #0 1	R	625	BRIDGES & MILLER LIMESTONE SCRUBBING	COAL 5.2 PERCENT SULFUR	6/73
KANSAS CITY POWER & LIGHT HEATHEN #0 4	R	100	COMBUSTION ENGINEERING LIMESTONE INJECTION & AET SCRUB	COAL 0.6- 3.0 PERCENT SULFUR	8/72
KANSAS CITY POWER & LIGHT HEATHEN #0 3	R	140	COMBUSTION ENGINEERING LIMESTONE INJECTION & AET SCRUB	COAL 0.6- 3.0 PERCENT SULFUR	11/72
KANSAS POWER & LIGHT LAWRENCE #0 5	R	400	COMBUSTION ENGINEERING LIMESTONE INJECTION & AET SCRUB	COAL 3.5 PERCENT SULFUR	11/71
KANSAS POWER & LIGHT LAWRENCE #0 4	R	125	COMBUSTION ENGINEERING LIMESTONE INJECTION & AET SCRUB	COAL 3.5 PERCENT SULFUR	12/68
LOUISVILLE GAS & ELECTRIC PADUCAH #0 6	R	65	COMBUSTION ENGINEERING LINE SCRUBBING	COAL 3.5-4.0 PERCENT SULFUR	4/73
NEVADA POWER RENU GARDNER #0 1	R	125	COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING	COAL 0.5- 1.0 PERCENT SULFUR	12/73

TABLE II
OPTIMIZED FGD SYSTEMS AS OF 10/79

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	PROCESS/VENDOR	FULL CHARACTERISTICS	MO/YR
LEVITT PURCH RIO GRANDE RD 2	N	125	CONFESSION EQUIP. ASSOCIATES SOLAR CARBONATE SCREENING	COAL 0.5- 1.0 PERCENT SULFUR	12/73
POTOMAC ELECTRIC & POWER WICKESBURG RD 2	N	100	CHENICO FARMINGTON OXIDE SCREENING	COAL 2.0 PERCENT SULFUR	9/73
SOUTHERN CALIFORNIA Edison MCRAVE RD 2	N	160	SLE/ STERN, ROGER LINE SCREENING	COAL 0.5- 0.8 PERCENT SULFUR	11/73
TEKESSEE VALLEY AUTHORITY SPENCER RD. 10	N	30	TEKESSEE VALLEY AUTHORITY LIME/LIMESTONE SCREENING	COAL	4/72

TABLE 12
FGD SYSTEMS UNDER CONSTRUCTION AS OF 10/77

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	PROCESS/VENDOR	FUEL CHARACTERISTICS	MO/YR
CENTRAL ILLINOIS LIGHT CO. DUCK CREEK NO. 1	N	100	KILLY STORLEY/ENVIRONMENTAL LIMESTONE SCRUBBING	COAL 2.5-3.0 PERCENT SULFUR	3/76
DETROIT Edison ST. CLAIR NO. 6	N	180	PEARSON ENGINEERING LIMESTONE SCRUBBING	COAL 3.7 PERCENT SULFUR	11/74
CONQUEST LIGHT ELBANA	N	510	CHENILO LIME SCRUBBING	COAL	5/75
KENTUCKY UTILITIES GREEN RIVER 1 & 2	N	64	AMERICAN AIR FILTER LIME SCRUBBING	COAL 3.8 PERCENT SULFUR	4/75
PEPPER POWER COLSTRIP NO. 2	N	360	COMBUSTION EQUIP. ASSOCIATES LIME SCRUBBING	COAL 0.8 PERCENT SULFUR	5/76
MONTANA POWER COLSTRIP NO. 1	N	360	COMBUSTION EQUIP. ASSOCIATES LIME SCRUBBING	COAL 0.8 PERCENT SULFUR	5/75
NEWARK POWER RELO GAS NO. 3	N	125	COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING	COAL 0.5-1.0 PERCENT SULFUR	6/75
INDIANA POWER SERVICE NATIONAL NO. 11	N	115	DAY POWERGAS ALLIANT LOR	COAL 3.2-3.5 PERCENT SULFUR	6/75
KENTUCKY STATE'S POWER SHERBOUR NO. 1	N	600	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL 1.0 PERCENT SULFUR	5/76
KENTUCKY STATE'S POWER SHERBOUR NO. 2	N	600	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL 1.0 PERCENT SULFUR	3/77
PENNSYLVANIA POWER CO. BRUCE MAUSFIELD NO. 2	N	600	CHENILO LIME SCRUBBING	COAL 4.3 PERCENT SULFUR	8/75
PENNSYLVANIA POWER CO. BRUCE MAUSFIELD NO. 1	N	600	CHENILO LIME SCRUBBING	COAL 4.3 PERCENT SULFUR	6/75
PENNSYLVANIA ELECTRIC EDDYSTONE NO. 1	N	120	UNITED ENGINEERS PACALON OXIDE SCRUBBING	COAL 2.5 PERCENT SULFUR	12/74
PUBLIC SERVICE OF INDIANA GINSCH NO. 2	N	650	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 1.5 PERCENT SULFUR	5/75
SOUTHERN CALIFORNIA Edison MORAVE NO. 1	N	160	UNIVERSAL OIL PRODUCTS LIMESTONE SCRUBBING	COAL 0.5-0.8 PERCENT SULFUR	12/74

TABLE 12

FGW SYSTEMS UNDER CONSTRUCTION AS OF 10/74

UTILITY COMPANY POWER STATION	NAME OR METROFIT	SIZE OF FGW UNIT (MW)	PROCESS/VENTION	FUEL CHARACTERISTICS	MO/YR
SOUTHWESTERN PUBLIC SERVICE HARRINGTON NO 1	N	345	COMBUSTION ENGINEERING LIME/LIMESTONE SCRAMBLING	COAL 0.5 PERCENT SULFUR	8/76
SPRINGFIELD UTILITY BOARD SOUTHWEST NO 1	N	200	UNIVERSAL OIL PRODUCTS LIMESTONE SCRAMBLING	COAL	6/76
TENNESSEE VALLEY AUTHORITY WIDOWS CREEK NO A	N	550	TENNESSEE VALLEY AUTHORITY LIMESTONE SCRAMBLING	COAL 3.7 PERCENT SULFUR	1/77

TABLE 13
PLANNED FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	VENDOR/PROCESS	FUEL CHARACTERISTICS	MO/YR
ARIZONA PUBLIC SERVICE CHULA VILLO 3	N	250	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	6/78
CENTRAL ILLINOIS PUBLIC SERV HEWTON NO.1	N	600	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 2.6-3.2 PERCENT SULFUR	3/77
INDIANAPOLIS POWER AND LIGHT PETERSBURG NO. 3	N	515	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 3.0-3.5 PERCENT SULFUR	4/77
KANSAS POWER & LIGHT JEFFERY NO. 2	N	700	COMBUSTION ENGINEERING LIMESTONE SCRUBBING	COAL	6/79
SOUTHERN MISSISSIPPI PWR COOP MATTESBURG NO.2	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
S. CAROLINA PWR SERV AUTHORITY GEORGETOWN NO. 2	N	140	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 1.0 PERCENT SULFUR	5/77
TEXAS UTILITIES MARTIN LAKE NO.4	N	795	RESEARCH COTTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	12/79
TEXAS UTILITIES MARTIN LAKE NO.3	N	795	RESEARCH COTTRELL LIMESTONE SCRUBBING	COAL 0.44 PERCENT SULFUR	12/78
SOUTHERN MISSISSIPPI PWR COOP MATTESBURG NO.1	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
TEXAS UTILITIES MONTICELLO NO. 3	N	600	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.8- 1.0 PERCENT SULFUR	12/78
CONSISTENT FGD SYSTEM					
LOUISVILLE GAS & ELECTRIC CANE RUN NO. 1	N	110	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/80
COLORADO UTE ELECTRIC HAYDEN NO. 2	N	250	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
COLORADO UTE ELECTRIC CHATEL STATION NO.1	N	450	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
COLORADO UTE ELECTRIC CRAIG STATION NO.2	N	450	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
CINCINNATI GAS AND ELECTRIC EAST BECK NO. 1	N	600	NOT SELECTED NOT SELECTED	COAL 0.5- 0.8 PERCENT SULFUR	6/79

TABLE 13
PLANTED FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETROFIT	SIZE OF FGD UNIT (MW)	VELOC/PROCESS	FUEL CHARACTERISTICS	MO/YR
ALABAMA ELECTRIC COOP BIG LEE NO 3	N	225	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.8- 1.5 PERCENT SULFUR	1/79
BASIN ELECTRIC MISSOURI BASIN NO 2	N	550	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/80
BASIN ELECTRIC MISSOURI BASIN NO 3	N	550	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/83
BASIN ELECTRIC MISSOURI BASIN NO 1	N	550	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 4	N	425	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/79
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 1	N	330	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/79
LOUISVILLE GAS & ELECTRIC CANE RUN NO 6	N	277	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/77
LOUISVILLE GAS & ELECTRIC CANE RUN NO 5	N	163	NOT SELECTED LIME SCRUBBING	COAL 3.5-4.05 PERCENT SULFUR	6/76
LOUISVILLE GAS & ELECTRIC MILL CREEK NO 2	N	330	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/78
LOUISVILLE GAS & ELECTRIC CANE RUN NO 2	N	107	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/80
ARIZONA PUBLIC SERVICE CHULA NO. 4	N	372	NOT SELECTED LIMESTONE SCRUBBING	COAL	6/79
PENNSYLVANIA POWER CO. BRUCE WATSFELD NO. 5	N	800	NOT SELECTED LIME SCRUBBING	COAL 4.3 PERCENT SULFUR	8/76
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 2	N	175	CHEMICO NOT SELECTED	COAL 0.8 PERCENT SULFUR	12/76
ARIZONA PUBLIC SERVICE FOUR CORNERS NO 3	N	229	CHEMICO LIME SCRUBBING	COAL 0.8 PERCENT SULFUR	3/77
NORTHERN INDIANA PUB SERVICE BAILY NO. 7	N	190	NOT SELECTED NOT SELECTED	COAL 3 PERCENT SULFUR	0/ 0

TABLE 12
PLANTED FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	AREA OR RETROFIT	SIZE OF FGD UNIT (MW)	VERSION/PROCESS	FUEL CHARACTERISTICS	MO/YR
NORTHERN INDIANA P&S SERVICE BAILY NO. 8	R	400	NOT SELECTED NOT SELECTED	COAL 3 PERCENT SULFUR	0/0
ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 4	R	600	CHERIL LIME SCRUBBING	COAL 0.9 PERCENT SULFUR	4/77
COLORADO UTE ELECTRIC HAYDEN NO. 1	R	202	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.45 PERCENT SULFUR	6/78
TEX. ELECTRIC ELEC SYSTEM IRVING NO. 3	R	650	NOT SELECTED NOT SELECTED	COAL 0.3 PERCENT SULFUR	0/0
ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 1	R	175	CHERIL LIME SCRUBBING	COAL 0.6 PERCENT SULFUR	10/76
MONTANA POWER COLSTRIP NO. 3	R	700	NOT SELECTED NOT SELECTED	COAL 0.7 PERCENT SULFUR	7/78
LOUISVILLE GAS & ELECTRIC C&E NO. 3	R	137	NOT SELECTED LIME SCRUBBING	COAL 3.5- 4.0 PERCENT SULFUR	6/80
MONTANA POWER COLSTRIP NO. 4	R	700	NOT SELECTED NOT SELECTED	COAL 0.7 PERCENT SULFUR	0/79
ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 5	R	600	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 0.6 PERCENT SULFUR	6/77
ALABAMA ELECTRIC CORP. LIGONIE NO. 2	R	225	NOT SELECTED LIMESTONE SCRUBBING	COAL 0.6- 1.5 PERCENT SULFUR	1/79
PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 3	R	500	NOT SELECTED NOT SELECTED	COAL 0.8 PERCENT SULFUR	5/78
PUBLIC SERVICE OF NEW MEXICO SAN JUAN NO. 4	R	500	NOT SELECTED NOT SELECTED	COAL 0.8 PERCENT SULFUR	5/80
SALT RIVER PROJECT NAVAJO NO. 1	R	750	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	3/76
SOUTHERN CALIFORNIA Edison POMONA NO. 1	R	600	NOT SELECTED LIME SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	6/77
SALT RIVER PROJECT NAVAJO NO. 3	R	750	NOT SELECTED LIME/LIMESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	3/77

TABLE 13
PLANTED FGD SYSTEMS AS OF 10/74

UTILITY COMPANY POWER STATION	NEW OR RETRIFIED	SIZE OF FGD UNIT (MW)	VENTILATION/PROCESS	FUEL CHARACTERISTICS	MO/YR
SALT RIVER PROJECT NAVAJO NO 2	1	750	NOT SELECTED LIME/LINESTONE SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	10/76
SOUTHERN CALIFORNIA Edison PORTAGE NO 2	R	640	NOT SELECTED LIME SCRUBBING	COAL 0.5- 0.8 PERCENT SULFUR	6/77

STATUS OF FLUE GAS DESULFURIZATION TECHNOLOGY IN JAPAN

Jumpei Ando

Faculty of Science and Engineering

Chuo University

Kasuga, Bunkyo-ku, Tokyo

The regulations on SO_2 emission have become increasingly stringent in Japan, forcing industry to install flue gas desulfurization (FGD) units. Major power companies have decided to install full-scale FGD plants for utility boilers with capacities larger than 300MW. All of the large plants will use wet processes by-producing gypsum. The present paper will comprehensively describe wet lime-limestone processes and double alkali FGD processes. In addition, three new processes, the Mitsui Miike limestone process, the Kurabo ammonium sulfate lime process, and the Dowa aluminum sulfate limestone process will be described in some detail.

STATUS OF FLUE GAS DESULFURIZATION TECHNOLOGY IN JAPAN

I Foreword

The oil crisis of late 1973 and the serious inflation resulting from it has affected considerably Japan's energy and desulfurization policies. Efforts more strenuous than before are to be made for the development of atomic energy and the import of LNG and coal. Imported crude oil which has supplied more than 70% of the total energy needs of Japan and has been the major source of SO₂ emission is expected to continue increasing at an annual rate of 7%, as against the 20-30% in the past. On the other hand, construction of many plants for hydrodesulfurization and gasification desulfurization of oil to reduce sulfur contents to below 0.3% which was planned eagerly in early 1973 was postponed or given up because of the inflation—the investment cost which was about 30 million dollars for a plant has nearly doubled.

The regulations on SO₂ emission have become even more stringent and have forced industry to install flue gas desulfurizers. Major power companies have finally decided to install full-scale SO₂ removal plants for utility boilers with capacities larger than 300MW. All of the big plants are to use wet processes by-producing gypsum because the reliability of the processes has been well demonstrated and the demand for gypsum is growing. Most of the large units are to be located outside city limits, because even 90% removal of SO₂ may not be satisfactory in the future in urban areas. Tokyo Electric Power, for example, has started to burn a considerable amount of LNG which gives off no SO₂ although it is expensive. Another reason for using LNG is the low NO_x emission involved. NO_x has aroused serious concern in Japan recently because of photochemical smog which has plagued big cities for the last few years. Simultaneous removal of SO₂ and NO_x has been enthusiastically studied by many organizations although there seems to be much difficulty involved in the process.

II Introduction

At the FGD symposium of EPA last year, the author reviewed seven major processes in Japan, namely those of Mitsubishi-JECCO, Chemico-Mitsui, Chiyoda, Wellman-MKK, Kureha-Kawasaki, Showa Denko and NKK.¹⁾ The present paper will describe comprehensively the technology and problems of wet lime-limestone processes, double alkali type processes and others. Some detailed description will be given on three new processes which might be of interest for application in the U.S.A., namely, the Mitsui Miike limestone process, the Kurabo ammonium sulfate lime process, and the Dowa aluminum sulfate limestone process.

III Wet lime-limestone process

SO₂ removal plants using wet lime-limestone processes with a capacity larger than 20MW equivalent are listed in Table 1. The Mitsubishi-JECCO process has been used most widely for oil-fired boilers, iron-ore sintering plants, etc. while the Chemico Mitsui and Mitsui Miike processes have been applied to coal-fired boilers. Five other processes have also been used mainly for flue gas from oil-fired boilers. Many of the plants use lime to obtain a high SO₂ removal efficiency—more than 90%—which is required in many districts in Japan. Limestone scrubbing removes 85 to 90% of the SO₂ at 0.9 to 1.2 stoichiometry by using a scrubber nearly twice as tall as that using lime.

A flow sheet common to most processes except the Chemico-Mitsui and the Mitsui Miike processes is shown in Figure 1. Types of scrubbers and examples of operation parameters are listed in Table 2.

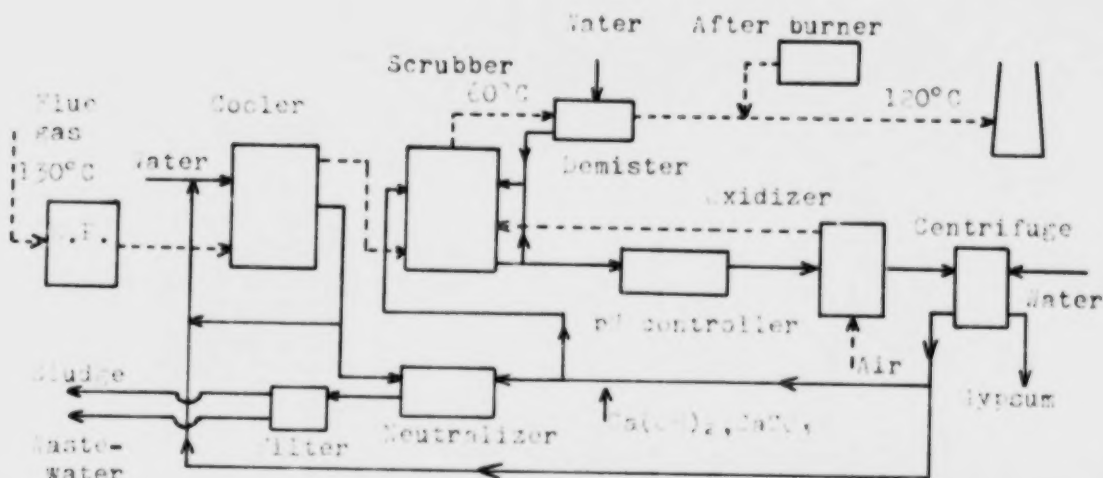


Figure 1. A Schematic of the wet lime-limestone process.

Cooler Flue gas from the electrostatic precipitator is first led into the cooler or prescrubber where the gas is sprayed with water. The cooler, which is not usually used in the United States, has two functions. (1) Cooling and humidifying the gas in order to aid in the prevention of caking in the scrubber. (2) Removal of dust and other impurities in the gas which were not caught by the electrostatic precipitator. This is useful in obtaining high purity by-products with good commercial value.

Table 1 Sulfur dioxide scrubbing installations in Japan (lime-limestone scrubbing)

Process developer	User	Plant site	Absorbent	MW ^a	Type of plant ^b	Year of completion	Gypsum (tons/day)
Mitsubishi-JECCO	Kansai Electric	Amagasaki	Ca(OH) ₂	30	Utility boiler	1972	20
"	Onahama Smelter	Onahama	"	29	Copper smelter	"	450
"	Kawasaki Steel	Chiba	"	37	Sintering plant	1973	16
"	Kansai Electric	Kainan	"	150	Utility boiler	1974	22
"	Toboku Electric	Hachinoe	"	125	"	"	53
"	Tokyo Electric	Yokosuka	CaCO ₃	130	"	"	20
"	Kyushu Electric	Shinkanda	Ca(OH) ₂	175	"	"	77
"	Kawasaki Steel	Mizushima	"	232	Sintering plant	"	130
"	"	"	"	279	"	1975	100
"	"	Chiba	"	130	"	"	30
"	Kansai Electric	Amagasaki	"	125	"	"	42
"	Teijin	Ehime	"	83	Industrial boiler	"	93
"	Mizushima Power	Mizushima	"	192	Utility boiler	"	105
"	Niigata Power	Niigata	CaCO ₃	117	"	"	47
"	Kyushu Electric	Katsura	"	250	"	"	75
"	"	"	"	175	"	1976	50
"	"	Ainoura	"	250	"	"	105
"	"	"	"	250	"	"	105
"	Chugoka Electric	Oase	Ca(OH) ₂	375	"	"	365
"	"	"	"	375	"	"	365
Chemico-Mitsui	Mitsui Aluminum	Omuta	"	119 ^c	Industrial boiler	1972	^d
Mitsui Mike	"	"	CaCO ₃	25 ^c	"	1974	22
"	"	"	"	175 ^c	"	1975	220
"	Elec. Pow. Dev.	Takasago	"	250 ^c	Utility boiler	1974	220
Bahco-Tsukishima	Yabagi Iron	Nagoya	Ca(OH) ₂	26	Sintering plant	1971	28
Babcock-Hitachi	Chugoku Electric	Mizushima	CaCO ₃	104	Utility boiler	1974	28
"	"	Tamashima	"	500	"	1975	385
Ishikawajima-TCA	Chichibu Cement	Kumagaya	Ca(OH) ₂	62	Diesel engine	1972	24
Sumitomo-Fuji Kasui	Sumitomo Metal	Kokura	"	32	Heating furnace	1974	28
Chubu-MKK	Ishihara Sangyo	Yokkaichi	CaCO ₃	77	Industrial boiler	"	55
"	Mitsubishi Gas	"	Ca(OH) ₂	22	"	"	13

^a Actual for boilers and equivalent gas flow for others (at 1900scfm per MW).^b Boilers are oil-fired unless noted.^c Coal-fired.^d Waste sludge of calcium sulfite.

Table 2 Example of operation parameters of FGD plants by-producing gypsum and calcium sulfite

Process developer	Absorbent, precipitant (stoichiometry)	Capacity, city 1,000scfm	Type of absorber	Slurry or solution pH conc. %	L/G gal/1,000scf	Space velocity ft/sec	Pressure drop inches	SO ₂ ppm in	SO ₂ ppm out	Moisture % of gypsum
<u>Wet lime-limestone process</u>										
Mitsubishi-JECCO	Ca(OH) ₂ 0.9-1	210	GP ^a	15	30-50	10	6	1,000	80	10
"	CaCO ₃ 0.9-1	230	GP	5-6	12	50-70	5	250	20	8-10
Chemico-Mitsui	Ca(OH) ₂ 1-1.05	210	Venturi	7	3-5	80-100	16	2,000	200	
Mitsui Milke	CaCO ₃ 0.9-1	44	Venturi	6	5	80-100		2,000	200	10-15
Babcock-Hitachi	CaCO ₃ 1-1.2	180	ppb	6.2	7-8	50	20	500	60	8-10
Chubu-MKK	CaCO ₃	150	Screen	6	10	70	3	1,500	200	10-12
Ishikawajima	Ca(OH) ₂	60	TCA	2	50	10		700	50	10-15
Sumitomo-Fuji	Ca(OH) ₂ 1-1.2	37	ppb	6	10	70	18	1,000	50	10-12
<u>Indirect lime-limestone process</u>										
Kureha-Kawasaki	Na ₂ SO ₃ , CaCO ₃	230	GP ^a	7	20	7-15	8	800	20	5-6
Showa Denko	Na ₂ SO ₃ , CaCO ₃	255	Cone	6.8	25	7-15		1,400	40	7-8
Nippon Kokan	(NH ₄) ₂ SO ₃ , CaCO ₃	90	Screen	6	30	14	10	700	30	10
Chiyoda	dil. H ₂ SO ₄ , CaCO ₃	466	Tellerette 1	2-4	200-300	3		600	50	6-8
Kurabo	(NH ₄) ₂ SO ₄ , Ca(OH) ₂	3	Tellerette 3-4	10	40-70	6	2	1,500	80	8-10
Dowa	Al ₂ (SO ₄) ₃ , CaCO ₃	82	Tellerette 3-4	10	20-60	4	4	600	20	10-12
Hitachi-Tokyo Elec.	Carbon, CaCO ₃	250	Packed			1.5	19	500	80	10-12

a) Grid packed

b) Perforated plate

Pressure drop includes that of absorber and mist eliminator

Scrubber The type of scrubber is a plastic grid packed tower for Mitsubishi-JECCO, a venturi for Chemico-Mitsui and Mitsui Miike, a TCA with polyethylene balls for Ishikawajima, a Bahco type for Tsukishima, a stainless steel screen type for Chubu-Mitsubishi Chubu-MKK (CM), and perforated plates for Babcock-Hitachi and Sumitomo-Fuji Kasui processes. The reliability of the Mitsubishi-JECCO and Chemico scrubbers has been well demonstrated in Japan. The former has less pressure drop. The TCA scrubber has a high SO₂ removal efficiency but presents the problem of wearing of balls. The Bahco and screen type scrubbers might be more susceptible to scaling. The perforated plate scrubber for the Sumitomo-Fuji Kasui process (Moretana type) is designed to give extreme turbulence producing foam layers 15-20 inches thick; a high SO₂ removal efficiency is attained while heavy mist is formed.

For the Chemico-Mitsui and Mitsui Miike processes which use no cooler, the scrubber is operated with a dilute slurry at a large L/G ratio to prevent scaling. The by-product contains a considerable amount of impurities as will be described in the author's other paper at the present symposium.²⁾

Water balance In most plants, the mist eliminator and by-product gypsum are washed with fresh water. The wash water is then fed mainly into the scrubber system and partly into the cooler. Some fresh water is also fed into the cooler. The amount of total input water more or less exceeds the output, i.e. water volatilized and taken into gypsum as water of crystallization and moisture. Therefore, a portion of an acidic liquor from the cooler is sent into the water-treatment system, neutralized with lime to precipitate heavy metals and other impurities, filtered, and then purged. The water balance is thus maintained. The use of fresh water and the purge of some wastewater help in the prevention of caking and corrosion.

It would be possible to apply a closed loop which would release no water from the system and yet not cause scaling. The accumulation of chlorine in the system, however, would cause corrosion problems. Also, the use of highly resistant materials would add considerably to the cost. Tohoku Electric Power Co. first planned to use a closed system for a 125MW plant based on the Mitsubishi-JECCO process, but the investment cost increases with the closed system; Tohoku finally changed the plan so that some water is purged after being treated.

Oxidation The rotary atomizer invented by JECCO has been used for the Mitsubishi-JECCO, Babcock-Hitachi, and Chubu-MKK processes because of its high efficiency and reliability. Different types of oxidizers have been developed for other processes. Low pH (3.5-4.0) is desirable for oxidation of calcium sulfite slurry. This is attained with a pH controller by adding a small amount of sulfuric acid as is being done at many plants, or by using flue gas and a catalyst as in the Mitsui Miike process. The facilities to produce gypsum, pH controller, oxidizer and centrifuge, occupy about 40% of the total investment cost. Nevertheless, the production of gypsum has been growing in Japan. Mitsui Aluminum Co.

Table 3 Indirect lime-limestone processes (double alkali type)

Process developer	Absorbent, precipitant	User	Plant site	M ^a	Type of plant	Year of com- pletion	Gypsum (tons/day)
Nippon Kokan	(NH ₄) ₂ SO ₃ , Ca(OH) ₂	Nippon Kokan	Keihin	46	Sintering plant	1972	17
Chiyoda	dil. H ₂ SO ₄ , CaCO ₃	Fuji Kasei	Kainan	50	Industrial boiler	"	23
"	"	Mitsubishi Rayon	Otake	27	"	1973	23
"	"	Daicel	Aboshi	31	"	"	25
"	"	Mitsubishi Chem.	Yokkaichi	130	"	1974	64
"	"	Mitsubishi Pet. Chem.	"	230	"	"	39
"	"	Hokuriku Electric	Toyama	250	Utility boiler	"	205
"	"	"	Fukui	350	"	1975	320
"	"	Denki Kagaku	Chiba	37	Industrial boiler	"	37
Kureha-Kawasaki	Na ₂ SO ₃ , CaCO ₃	Tohoku Electric	Shinsendai	150	Utility boiler	1974	44
"	"	Shikoku Electric	Shintokushima	450	"	1975	330
"	"	"	Sakaide	450	"	"	330
Showa Denko	"	Showa Denko	Chiba	150	Industrial boiler	1973	120
"	"	Showa Pet. Chem.	Kawasaki	62	"	1974	77
"	"	Kanagafuchi Chem.	Takasago	93	"	"	90
Showa Denko-Ebara	"	Nippon Mining	Saganoseki	37	H ₂ SO ₄ plant	1973	CaSO ₃
"	"	Yokohama Rubber	Hiratsuka	27	Industrial boiler	1974	22
"	"	Nisehin Oil	Isogo	27	"	"	33
"	"	Poly Plastic	Fuji	65	"	"	77
"	"	Kyowa Pet. Chem.	Yokkaichi	46	"	"	38
"	"	Toho Zinc	Annaka	43	"	"	77
"	"	Ajinomoto	Yokkaichi	25	H ₂ SO ₄ plant	"	28
Tsukishima	Na ₂ SO ₃ , Ca(OH) ₂	Kinura Utility	Nagoya	63	Industrial boiler	"	45
"	"	Dai-ikuwa Paper	Fuji	85	"	1975	50
Kurabo Engineering	(NH ₄) ₂ SO ₄ , Ca(OH) ₂	Kurarey	"	31	"	1974	33
"	"	Daicel	Aboshi	53	"	1975	44
Dowa Mining	Al ₂ (SO ₄) ₃ , CaCO ₃	Dowa Mining	Okayama	82	H ₂ SO ₄ plant	1974	26
Hitachi-Tokyo Electric	Carbon, CaCO ₃	Tokyo Electric	Kashima	150	Utility boiler	1972	55

^a Actual for boilers and equivalent gas flow for others. Boilers are oil-fired.

which has been producing waste calcium sulfite sludge will also change the process to by-produce gypsum as will be described later. By-production of calcium sulfite sludge and its stabilization as it is being done in the United States might be of much interest in the future in Japan when there is an oversupply of gypsum.

IV Indirect lime-limestone process (double alkali type)

Many double alkali type processes that use lime or limestone as a precipitant have been developed in Japan including those which use an acidic solution or acid as the absorbent. The Hitachi-Tokyo Electric process uses activated carbon as the absorbent and limestone as the precipitant. All of those processes are classified in the category of "indirect lime-limestone process". The installations with a capacity larger than 20MW equivalent are listed in Table 3. The operation parameters are shown in Table 2.

Wet process The various processes use different absorbents; a sodium sulfite solution is used for the Showa Denko, Kureha-Kawasaki, and Tsukishima processes, an ammonium sulfite solution for the NKK process, an acidic ammonium sulfate solution for the Kurabo process, an aluminum sulfate solution for the Dowa process, and a dilute sulfuric acid with iron sulfate for the Chiyoda process. The pH of the solution is 6 to 7 for ammonium and sodium sulfites, 3 to 4 for ammonium and aluminum sulfates and 1 for sulfuric acid. The L/G ratio is 7-15 (gal/1,000scf) for the solutions with pH 6-7, 20-70 for the solutions with pH 3-4, and 200-400 for the acid at pH 1. The more acidic the solution is, the less the SO₂ absorption capacity, the lesser the problem of scaling, and the easier the reaction with limestone. Limestone can be reacted with a sodium bisulfite solution, as in the Showa Denko and Kureha-Kawasaki processes, but the reaction occurs slowly requiring large reaction vessels. Lime is used for the Tsukishima, NKK, and Kurabo processes.

For the Chiyoda, Dowa, and Kurabo processes, the liquors which absorbed SO₂ are contacted with air to oxidize SO₃⁻ into SO₄⁻. Limestone or lime is then added to precipitate gypsum. For other processes, limestone or lime is added first to precipitate calcium sulfite which is then oxidized into gypsum. By the double alkali type process, gypsum usually grows in larger crystals than with the wet lime-limestone process. Moisture content of the by-product gypsum after centrifugalization ranges from 6 to 12% as compared with 8-15% for the wet lime-limestone process.

The liquor from the gypsum centrifuge is returned mainly to the scrubber system. Softening of the liquor which is usually needed to prevent scaling, is not necessary when an acidic solution at a pH below 4 is used.

At most plants, a small portion of the liquor is purged to maintain the concentrations of chlorine, magnesium and other impurities under a certain level. Calcium sulfite obtained by the slow reaction of limestone and sodium bisulfite grows into fairly large crystals which are not difficult to handle.²⁾ On the other hand, lime reacts rapidly with a sodium bisulfite solution to give very fine crystals of calcium sulfite like those produced by the wet lime-limestone process.

Carbon absorption Another type of the indirect lime-limestone process is dry activated carbon absorption which is used in the Hitachi-Tokyo Electric process. The carbon which has absorbed SO_2 is washed with water to give a dilute sulfuric acid of 15-20% concentration. The acid is treated with powdered limestone to produce gypsum, which is centrifuged to 10-12% moisture. The process requires large absorption towers because low gas velocity (1.5 ft/sec) is used to obtain 85-90% removal of the SO_2 . Hitachi has recently operated a 1,700scfm pilot plant in which SO_2 is absorbed with a carbon slurry. The resulting sulfuric acid is treated with limestone to obtain gypsum.

V Other processes

There are about one hundred relatively small sodium scrubbing plants in operation by-producing mainly sodium sulfite and some sodium sulfate. Those processes were reviewed last year¹⁾ and there has not been much progress since then. Major plants using other processes are listed in Table 4.

Wellman-Lord process Many plants using the Wellman-Lord process have been constructed by Mitsubishi Chemical Machinery Co. (MCK) and also by Sumitomo Chemical Engineering Co. (SCEC). The reliability of the process has been well demonstrated. A main problem in the process is the oxidation of sodium sulfite into sulfate which does not absorb SO_2 . The sulfate has to be removed from the system, resulting in the loss of sodium and the need for wastewater treatment. At Nishinagoya Plant of Chubu Electric (220MW), 7 tons/day sodium hydroxide have been used for make-up giving the equivalent amount of sodium sulfate. SCEC uses an oxidation inhibitor to reduce the sulfate formation to below half. The inhibitor, however, might give some trouble in wastewater treatment by disturbing the oxidation of any reducing component in the water.

Magnesia scrubbing Two plants are in operation and one is under construction with different magnesium scrubbing processes, by Mitsui Mining, Onahama-Tsukishima, and Chemico-Mitsui. A common feature of these processes is the production of large $\text{MgSO}_4 \cdot 6\text{H}_2\text{O}$ crystals (200 microns or so) which are much easier to filter and dry than are the small $\text{MgSO}_4 \cdot 3\text{H}_2\text{O}$ crystals. Both the Mitsui Mining and Onahama plants have no problem in the filtration and drying steps which were troublesome at Boston Edison Co. A nearly completed Idemitsu Kosan plant based on the Chemico-

Table 4 Sulfur dioxide scrubbing installations in Japan which by-produce sulfuric acid and sulfur

Process developer	Absorbent	User	Plant site	M ^a	Type of plant	Year of completion	By-product (tons/day)
Wellman-MKK	Na ₂ SO ₃	Japan Synth. Rubber	Chiba	70	Industrial boiler	1971	H ₂ SO ₄ 44
"	"	Chubu Electric	Nishinagoya	220	Utility boiler	1973	" 88
"	"	Japan Synth. Rubber	Yokkaichi	150	Industrial boiler	1974	" 44
"	"	Toyo Rayon	Nagaya	103	"	"	" 50
"	"	Mitsubishi Chem.	Mizushima	186	"	"	" 88
"	"	Company K	Kawasaki	217	"	"	" 110
"	"	Kurashiki Rayon	Okayama	127	"	"	" 66
"	"	Company KKK	Kashima	400	"	"	" 120
"	"	Company SKD	Yokkaichi	124	"	"	" 60
Wellman-SCEC	"	Toa Nenryo	Kawasaki	23	Claus furnace	1971	S 55
"	"	Sumitomo Chiba Chem.	Chiba	120	Industrial boiler	1973	H ₂ SO ₄ 13
"	"	Fuji Film	Fuji	50	"	1974	Liquid SO ₂ 88
"	"	Sumitomo Chiba Chem.	Chiba	180	"	1975	H ₂ SO ₄ 25
"	"	Sumitomo Chem.	Niihama	50	"	"	" 240
Onahama-Tsukishima	MgO	Onahama Smelter	Onahama	28	Copper smelter	1972	" 18
Mitsui Mining	"	Mitsui Mining	Hibi	25	H ₂ SO ₄ plant	1971	" S
Chemico-Mitsui	"	Idemitsu Kosan	Chiba	162	Claus and boiler	1974	" 17
Sumitomo Shipbuilding	Carbon	Kansai Electric	Sakai	53	Utility boiler	1971	H ₂ SO ₄ S
Shell	CuO	Showa Yokkaichi	Yokkaichi	37	Industrial boiler	1974	S S
Mitsubishi-IFP	(NH ₄) ₂ SO ₃	Maruzen Oil	Shimozu	14	Claus furnace	"	S S
TEC-IFP	"	Fuji Oil	Chiba	3	"	"	" S

^a Actual for boilers and equivalent gas flow for others. Boilers are oil fired.

Mitsui process is to return the recovered SO_2 to a Claus furnace for sulfur production; the other plants by-produce sulfuric acid.

Dry carbon process By the Sumitomo Shipbuilding process, the SO_2 absorbed on activated carbon is expelled by heating it in a reducing gas, to release SO_2 gas of 10-20% concentration which is used for sulfuric acid production. Moving beds are used for both absorption and desorption. The operation has been carried out smoothly but there is no further plan to build larger plants, presumably because of the requirement of large absorbers and the considerable consumption of carbon.

Shell process In the Shell process, SO_2 is absorbed with copper oxide to form copper sulfate, which is then reduced with hydrogen to expel SO_2 . A commercial plant started operation several months ago at Yokkaichi. The process seems fairly costly requiring a large absorber and hydrogen although it has the advantage of a dry process which requires no reheating of treated gas.

Ammonia scrubbing-IFP process The Mitsubishi-IFP process uses ammonia scrubbing with the thermal decomposition of ammonium sulfite and sulfate to regenerate SO_2 , which is then reacted with H_2S in an IFP reactor to produce sulfur. A plant at Shimizu started operation several months ago. A similar plant has been recently constructed by Toyo Engineering for Fuji Oil. The process is not simple and may need further improvement before satisfactory operation is attained.

VI Mitsui Miike limestone gypsum process

Outline of process This process has been developed by Mitsui Miike Machinery Co. (1-1, 2-chome, Nihonbashi Muromachi, Chuo-ku, Tokyo). SO_2 is absorbed with a limestone slurry containing a catalyst which promotes the reaction. The resulting calcium sulfite slurry is oxidized into usable gypsum by further contact with flue gas and air. The catalyst also promotes the oxidation.

State of development Mitsui Miike constructed the desulfurization unit at the Omuta plant of Mitsui Aluminum Co. which has been operated successfully. The disposal of calcium sulfite sludge, however, is a considerable problem because of the poor nature of the sludge. In order to by-produce usable gypsum, Mitsui Miike has developed a new process based on a patent by Dr. S. Akimoto using a catalyst that promotes both the reaction of SO_2 and limestone and the oxidation of calcium sulfite. After tests with a pilot plant with a capacity to treat coal-fired flue gas at $2,500\text{Nm}^3/\text{hr}$ ($1,500\text{scfm}$), a larger test unit ($75,000\text{Nm}^3/\text{hr}$ of coal-fired flue gas) was completed in September 1974 at Omuta. A commercial plant at the Takasago Station of the Electric Power Development Co. with a capacity of treating $800,000\text{Nm}^3/\text{hr}$ flue gas from a coal-fired utility boiler (250MW) will be completed in late 1974. Mitsui Aluminum will install a $552,000\text{Nm}^3/\text{hr}$ unit (175MW) in 1975.

Process description A flowsheet of the process is shown in Figure 2. Flue gas is treated in two venturi scrubbers in series with a counter-current flow of limestone slurry containing a metallic catalyst. The resulting calcium sulfite slurry is then fed into a pH controller where the slurry contacts a portion of flue gas (5-10% of the total) to lower the pH and to complete the reaction of limestone with SO_2 . The slurry is then pumped into an oxidizing tower into which air is blown from the bottom to convert the sulfite into gypsum. The gypsum slurry is centrifuged. Most of the liquor from the centrifuge is used to prepare a limestone slurry. A small portion of the liquor is sent to a wastewater treatment system to prevent the accumulation of impurities. Gypsum grows into large crystals of 50-200 microns in size and can be used for wallboard and cement.

The catalyst prevents the formation of a calcium sulfate coating on calcium sulfite and carbonate and thus promotes both the reaction of the carbonate with SO_2 and the oxidation of the sulfite into sulfate as shown in Figure 3 and Table 5 which are based on the pilot tests. There is virtually no loss of catalyst into gypsum because gypsum can be washed well. In the above-mentioned wastewater treatment, the catalyst can be precipitated in the form of hydroxide by raising the pH and returning it to the absorbing system.

The capital required is about 50% higher for the lime-calcium sulfite sludge process. The pH controller and oxidizer add about 20% to the cost. If about 80% oxidation is satisfactory to improve the settling property in the disposal pond, as in the partial oxidation of the sulfite sludge, then the oxidizing system can be greatly simplified. The pond should be sealed to prevent the leakage of water containing the catalyst.

Table 5 Example of the composition of by-product (%)
(stoichiometry 0.95)

Absorbent	$\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$	CaSO_3	CaCO_3
CaCO_3 + Catalyst	98.2	0.7	trace
CaCO_3 only	71.0	15.3	12.3

Advantages High SO_2 recovery is attained with limestone slurry. Oxidation of calcium sulfite occurs rapidly. SO_2 in flue gas is used for pH adjustment instead of using sulfuric acid.

Disadvantage The facilities for oxidation add considerably to investment cost.

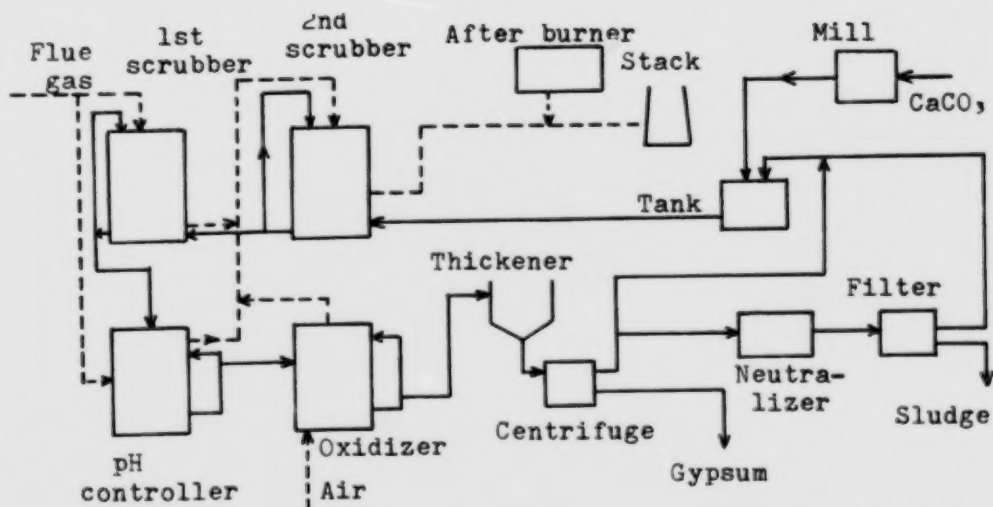


Figure 2 Flowsheet of Mitsui Miike limestone gypsum process

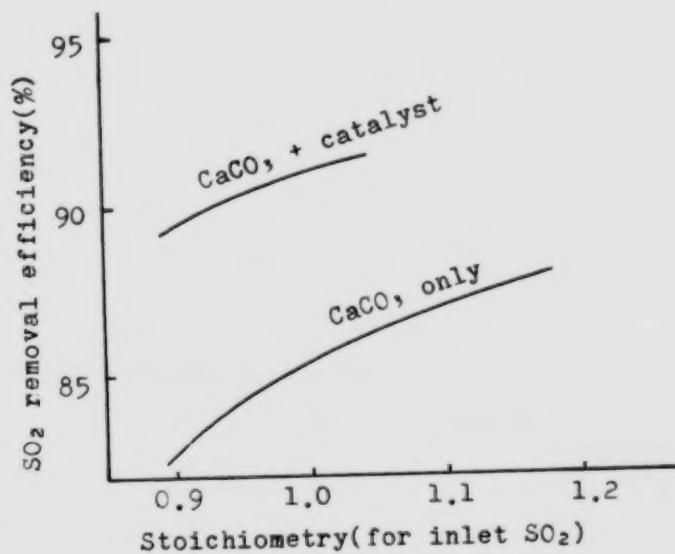


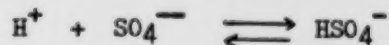
Figure 3 Effect of catalyst on SO₂ removal efficiency

VII Kurabo ammonium sulfate gypsum process⁴⁾

Outline This process has been developed by Kurabo Industries, Ltd. (2-41, Kitakyutaromachi, Higashi-ku, Osaka). SO₂ is absorbed by a slightly acidic ammonium sulfate solution at pH 3-4. The solution is then oxidized by air to give an acidic ammonium sulfate solution, which is then treated with lime to precipitate gypsum and to recover aqua ammonia which is returned to the absorbing system. Ammonium sulfate can be produced, if desired.

State of development Kurabo has built several small ammonia scrubbing plants in addition to many sodium scrubbing plants. Plume formation, the greatest problem for the ammonia process, has been fairly well solved by the use of a cold, dilute solution but this method may not be suitable to larger plants. Kurabo has recently developed a method to use an acidic ammonium sulfate solution as an absorbent to eliminate the plume and has successfully operated a 2MW equivalent test unit. Two commercial plants are presently under construction (Table 3).

Theory of absorption with ammonium sulfate solution Plume can be eliminated by the use of an acidic absorbing liquor because the vapor pressure of NH₃ is less than 1ppm with a solution at a pH lower than 4. The acidic ammonium sulfate solution has a larger capacity of SO₂ absorption than plain water or a saturated calcium sulfate solution as shown in Figure 4 because of the smaller pH drop due to the following equilibrium in the ammonium sulfate solution:



The minimum L/G ratio (liter/Nm³) required to remove 95% of the SO₂ is shown in Table 6. The relationship of L/G to SO₂ removal efficiency is shown in Figure 5. Basic data for the design of SO₂ absorbers obtained through the pilot tests are shown in Table 7.

Table 6 Inlet SO₂ concentration and minimum L/G
(gal/1,000scf, 60°C, 95% SO₂ removal)

<u>Absorbing liquor</u>		<u>Inlet SO₂ concentration, ppm</u>		
<u>Composition</u>	<u>Initial pH</u>	<u>1,000</u>	<u>2,000</u>	<u>3,000</u>
CaSO ₄ ·2H ₂ O - H ₂ O	7.0	98	148	191
0.25 mol/l (NH ₄) ₂ SO ₄	3.5	43	51	66
0.5 "	3.5	36	41	50
1.0 "	3.5	29	38	44

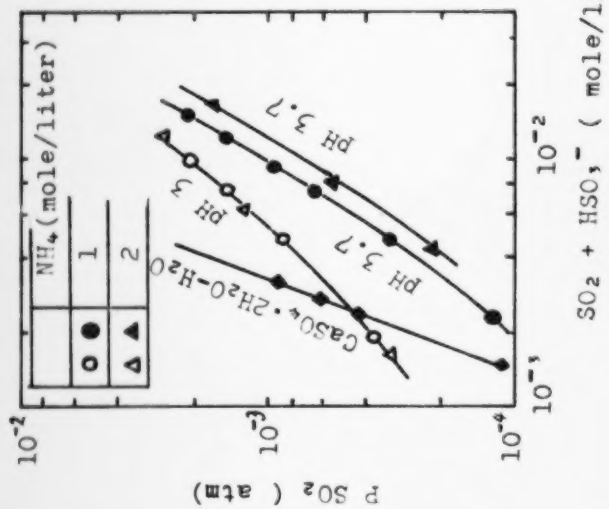


Figure 4 Relationship of $\text{SO}_2 + \text{HSO}_3^-$ (mole/liter) concentration to partial pressure of SO_2 (60°C)

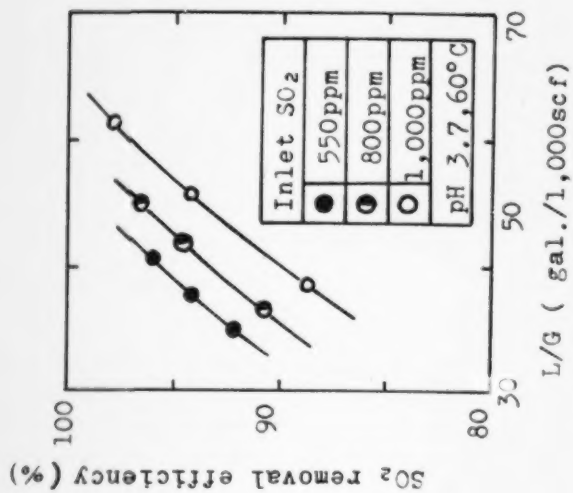


Figure 5 SO_2 removal efficiency at the test unit (2,950scfm)

Tower diameter 3.3ft
Height of packing 6.6ft
Packing Netring HA-18
Space velocity 4-5 ft/sec

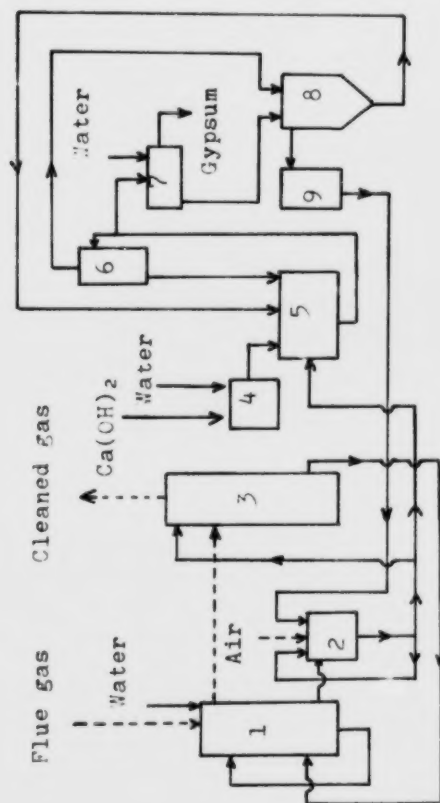
Table 7 Design base of SO₂ absorber

Packing	L/G (gal/1,000scf)	Pressure drop (inches Aq)	Space velocity (ft/sec)	Height of packing (ft)
Tellerette L type	56	4	6	12
Netring HA-18	56	4	4.5	9

Process description The flowsheet of the process is shown in Figure 6. Flue gas is first led into a KBCA scrubber and then into a packed tower absorber. The main function of the KBCA scrubber is to cool the gas to 60°C and to concentrate the absorbing liquor (ammonium sulfate solution). More than 90% of the SO₂ is removed. The gas is then reheated by the afterburning of oil. The liquor from the packed tower absorber is sent to the KBCA unit, concentrated, and then led into an oxidizer. The pH of the liquor in the oxidizer is adjusted to from 3 to 4 by adding dilute aqua ammonia and the sulfite in the liquor is oxidized into sulfate by small bubbles of air formed by introducing a jet stream of circulating liquor accompanying air into a pool of the liquor. About five times the stoichiometric amount of air is used. Tests have shown that Fe⁺⁺ catalyst gives an optimum oxidation rate at pH 4 and Mn⁺⁺ at pH 5 (Figure 7). Actually no catalyst is added for oxidation because it occurs fairly rapidly at a rate of 0.8-1.0 kg mole/m³/hr; presumably a small amount of vanadium and iron derived from the fuel helps the oxidation.

Most of the liquor from the oxidizer is returned to the absorber, and a portion is sent to a set of three reactors where the liquor is treated with milk of lime to precipitate gypsum. In order to raise the concentration of the slurry and to increase the retention time of gypsum for better crystal growth, a portion of the slurry is sent to a liquid cyclone. The gypsum is returned to the reactor and the liquor which contains some gypsum is sent to a thickener. The overflow from the thickener is sent to an aqua ammonia tank. The sludge from the thickener is returned to the reactor, and the other portion of the slurry is sent to a centrifuge. The liquor from the centrifuge and wash water are sent to an aqua ammonia tank, and the aqua ammonia is sent to the oxidizer.

The concentration of the absorbing liquor is normally about 0.5 mole of ammonium sulfate per liter at L/G 56. It can be raised to 2.5 moles at the same L/G ratio without appreciable decrease of SO₂ removal efficiency. The oxidation rate decreases slightly at the higher concentration. The amount of water added to the system, namely lime dissolving water and gypsum wash water, is usually less than that volatilized in the cooler. Therefore, some water is introduced into the cooler to maintain the water balance. Occasional purging may be required to prevent buildup of chloride. The by-product gypsum contains about 9% moisture with a bulk density of about 1.06 and pH 8. Average crystal size is 40 to 60 microns. The gypsum can be used for both wallboard and cement additives.



- | | |
|-----------------|---------------------|
| 1 KBCA scrubber | 6 Liquid cyclone |
| 2 Oxidizer | 7 Centrifuge |
| 3 Absorber | 8 Thickener |
| 4 Slaking tank | 9 Aqua ammonia tank |
| 5 Reactor | |

Figure 6 Flowsheet of Kurabo ammonium sulfate gypsum process

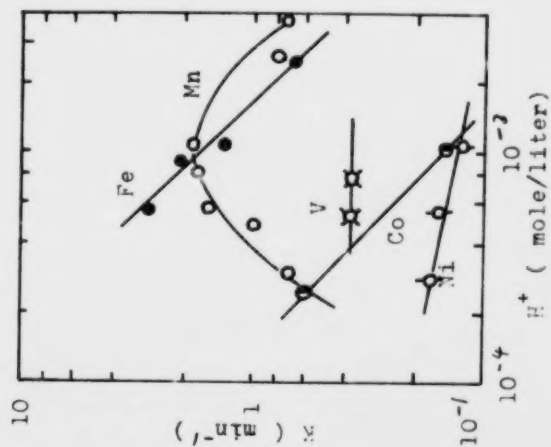


Figure 7 Oxidation rate of sulfite ion (40°C)
(HSO_3^+) $< 10^{-2}$ mole/liter
(M^{++}) 2×10^{-4} mole/liter

Ammonium sulfate can be obtained easily by concentrating the solution from the oxidizer. In this case, a nearly saturated solution (2.5-3 moles/liter) is used as the absorbent.

Cost Estimation An example of cost estimation is shown in Table 8.

Table 8 An example of cost estimation

Plant:	59,000 scfm
Oil consumption:	1,590,000 gal/year
Sulfur content:	2.8%
Operation hours:	8,500 hours/year
Construction cost:	\$4.0 million

<u>Material</u>	<u>Unit cost</u>	<u>Annual cost</u>
25% aqua ammonia	30.3 cents/gal.	6,000 dol.
Oil for reheating	22.7 "	112,400 "
Industrial water	7.6 "	1,680 "
Power	2.0 cents/kW	96,000 "
Quick lime	0.9 cents/lb.	63,000 "
Recovered gypsum	0.4 "	-114,600 "
Subtotal		162,800 "
Depreciation		515,000 "
Interest		160,000 "
Labor		24,000 "
Maintenance, etc.		64,000 "
Total		925,800 "
Desulfurization cost		5.9 cents/gal.oil

(1 dollar = 250 yen)

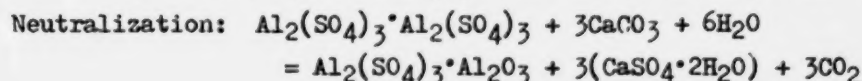
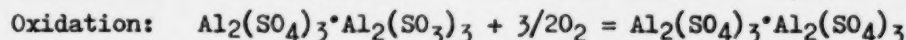
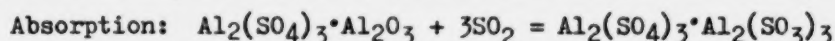
Advantages There is no plume and no scaling. Both oxidation and crystallization of gypsum are made at 50-60°C without external heating. The pH of liquor is kept over 3 to prevent corrosion. All steps are carried out at atmospheric pressure. A closed system can be achieved. Ammonium sulfate can be produced easily, if desired. The softening step of the circulating liquor is unneeded.

Disadvantages The process is less simple than the wet lime process. Limestone cannot be used.

VIII Dowa aluminum sulfate process

Outline of the process This process has been developed by Dowa Mining Co. (8-2, 1-chome, Marunouchi, Chiyoda-ku, Tokyo). Dowa is one of the biggest nonferrous metal manufacturers in Japan and owns many smelters and sulfuric acid plants. The aluminum sulfate process has been developed to desulfurize waste gas from smelters, roasters and sulfuric acid plants. The principle of the process is as follows:

SO₂ is absorbed in a solution of basic aluminum sulfate, Al₂(SO₄)₃·Al₂O₃, to form Al₂(SO₄)₃·Al₂(SO₃)₃. The liquor is oxidized by air into Al₂(SO₄)₃, which is then treated with powdered limestone to precipitate gypsum and to regenerate the basic aluminum sulfate solution.



State of development After tests with a pilot plant with a capacity of 170scfm, a small commercial plant (2,000scfm) was constructed at Taenaka Mining, Mobara Works and started operation in October 1972 to treat waste gas from a molybdenum sulfide roaster containing 7,500ppm SO₂ at 100°C. A larger pilot plant with a capacity of treating 1,700scfm tail gas from a sulfuric acid plant has also been in operation.

Two commercial units, each with a capacity of treating 140,000Nm³/hr (82,000scfm) tail gas from sulfuric acid plants, started operation in June 1974 at Okayama Works of Dowa. Mitsui Shipbuilding Co. has joined Dowa for further development of this process.

Absorbing liquor The relations between the composition of basic aluminum sulfate liquor (Table 9) and SO₂ absorbing capacity and the boiling point of the liquor after the SO₂ absorption are shown in Figures 8 and 9.

Table 9 Composition of solutions in Figures 8 and 9

Al ₂ O ₃ (grams/liter) in solution				
No.	Free	Combined	Total	Basicity (%)
I	34.4	66.1	100.5	34.2
II	40.0	58.2	98.2	40.8
III	45.5	53.2	98.7	46.1

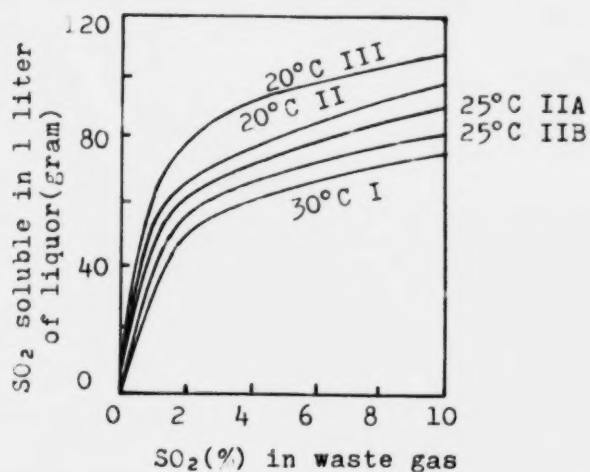


Figure 8 SO_2 absorbing capacity of the liquors

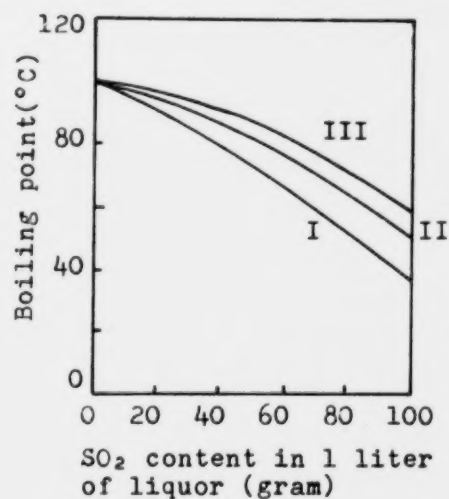


Figure 9 Boiling point of the liquor containing SO_2

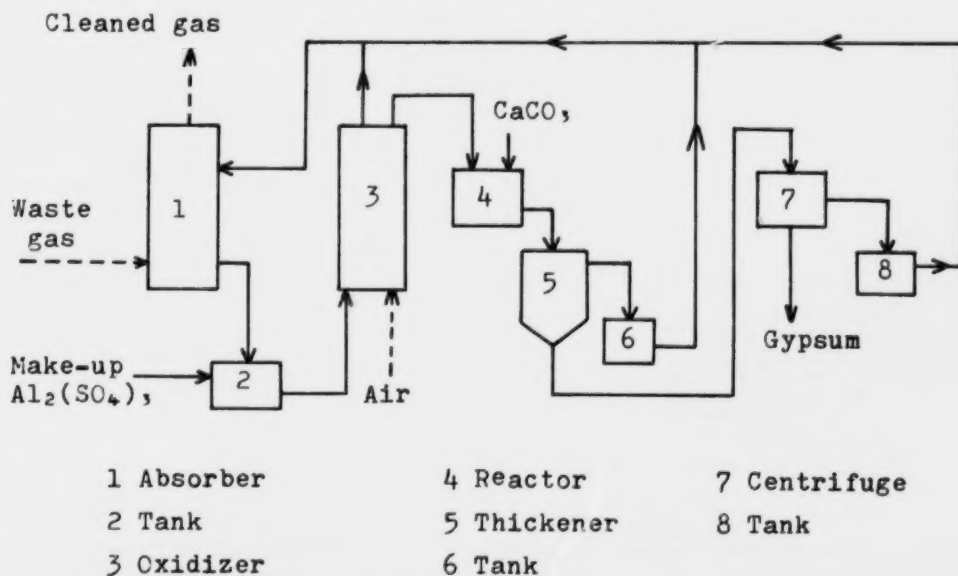


Figure 10 Flowsheet of Dowa aluminum sulfate limestone process

An optimum concentration as well as basicity of the absorbing liquor is selected according to the SO₂ concentration of gas and the removal efficiency required. Normally liquors at pH 3 to 4 which are more dilute than those shown in Table 9 are used.

Process description The flowsheet of the process is shown in Figure 10. For the small commercial plant of Taenaka Mining, a TCA scrubber is used to recover more than 90% of the SO₂ at L/G 35 (gal/1,000scf). For the pilot tests (1,700scfm) with tail gas at 85°C from a sulfuric acid plant, a packed tower (800mm in diameter) was used to recover more than 95% of the SO₂ at L/G 20. The liquor from the absorber is led into a tank to which make-up aluminum sulfate is added. The liquor from the tank is sent to an oxidizer where aluminum sulfite is oxidized into sulfate by small bubbles of air. The oxidizer, which has no moving parts, has been developed by Dowa. Twice the stoichiometric amount of air is used. Most of the oxidized solution is returned to the absorber, a portion is sent to a neutralizer and treated with powdered limestone (mostly under 200 mesh in size) under a condition suitable for the crystal growth of gypsum.

The slurry from the neutralizer passes through a thickener and is then centrifuged. All of the liquor from the thickener and the centrifuge as well as the wash water of gypsum are returned to the absorber. The chemical composition of the by-product gypsum from the pilot plant (1,700scfm) is shown in Table 10. A small amount of aluminum is contained in the gypsum. The consumption of aluminum is about 1 lb (as Al) for 1 ton of gypsum.

Table 10 Chemical composition of by-product gypsum (%)

<u>No.</u>	<u>CaO</u>	<u>SO₃</u>	<u>Al</u>	<u>Moisture</u>
1	32.80	46.65	0.05	10.0
2	32.34	46.43	0.05	12.1
3	32.03	46.61	0.04	12.0
4	32.24	45.68	0.06	10.8
5	32.88	45.86	0.05	12.8
6	32.07	46.48	0.05	11.8

Operation experiences The small commercial plant of Taenaka has been in operation since October 1973. There is no appreciable problem of scaling. Cleaning the pipings once every several months is enough to remove soft deposits. Continuous operation of the pilot plant (1,700scfm) for several months also produced no appreciable scaling. The operation of the new plants (82,000scfm) has also been satisfactory.

Advantages (1) Limestone is used for neutralization because of the low pH of the liquor. (2) The L/G ratio is relatively small. (3) The oxidation occurs fairly rapidly. (4) There is essentially no scaling and no wastewater. (5) A fully automatic operation can be achieved.

Disadvantage Aluminum sulfate is consumed although in a small amount.

IX Evaluation of the processes

Dry processes Now there are three dry processes, each being used for a prototype plant. Those are two activated carbon processes by Hitachi-Tokyo Electric and by Sumitomo Shipbuilding, and the Shell copper oxide process. In addition, pilot plants using the dry sodium absorption process followed by the decomposition of the resulting sodium sulfate has been operated by Tsukishima Kikai Co. and also by the National Research Institute of Pollution and Resources. There is no plan, however, to install a commercial plant using any of those dry processes. Mitsubishi Heavy Industries and Chubu Electric Power have recently given up the activated manganese process because of the difficulty of achieving 90% recovery and the economical disadvantage. The advantage of the dry process—no reheating of gas—may not compensate for the disadvantages—the need for a large absorber and an expensive absorbent or reducing agent.

Problems common to wet processes Numerous plants using wet processes are in operation or under construction. Still, the wet processes have some problems in common. One is reheating and the other is wastewater. For reheating of gas, afterburning of oil has been used exclusively in Japan. The oil consumption is about 5% of that for power generation, indicating that when the SO₂ removal efficiency by scrubbing is 90%, overall efficiency would drop to 85%. The addition of SO₂ and some dust to the gas is going to present another problem as the regulations on SO₂ and dust become more and more stringent. Studies on the use of the heat exchanger have been started recently.

It is common in Japan to purge some water from the system in order to maintain the concentration of impurities in the circulation liquor under a certain level. Among the impurities, chlorine derived from fuel and input water is most troublesome because it causes corrosion. Particularly when the multitube heat exchanger is used for reheating, the corrosion problem becomes significant. It may be more logical to purge some water after treating it thoroughly than to use expensive corrosion resistant materials. Where no water at all is allowed to be purged, some means of removing chlorine from the circulating liquor may be desired.

Wet lime-limestone processes Wet lime-limestone processes are most promising in Japan as in the United States. The reliability of the

Mitsubishi-JECCO process has been well demonstrated. However, most plants based on this process have treated gases with a low SO₂ concentration (300-1,000ppm) except for a plant of the Onahama Smelter which treats waste gas from a converter containing 25,000ppm SO₂. Two full-scale plants to treat flue gas with about 1,600ppm SO₂ from oil-fired utility boilers (375MW each) will start operation early in 1975. The operation performance should be watched for further evaluation. The Chemico-Mitsui process has also gained some fame from the successful operation of the Omuta plant of Mitsui Aluminum Co. using coal-fired flue gas. There has been much argument concerning the reason for the scale-free operation. Whether the circulation liquor is saturated or unsaturated with gypsum is a point which has not been made clear yet. However, there is no doubt that careful operation control is one of the keys to success. For example, the pH of the slurry is measured manually because an automatic pH meter might not be quite reliable.

The Mitsui Miike process has the advantages of attaining high SO₂ removal with limestone and of pH control without sulfuric acid. The Mitsui Miike process as well as the Mitsubishi-JECCO process may be useful also in the U.S.A. if there is any use for the by-product gypsum. The Mitsui Miike process may be applicable also for discarding the by-product, if the partial oxidation, which is required to improve the property of the sludge, can be done in a simple way without much cost.

For evaluation of other wet lime-limestone processes, a longer operation period or operation of larger plants may be needed.

Double alkali type processes Among the double alkali type processes, the Chiyoda process and sodium-limestone processes of Showa Denko and Kureha-Kawasaki have been used widely. Smooth operation and use of limestone may be the reason for the use. The Chiyoda process is simple and the plant operation is most easy, although it requires a big absorber and a high L/G ratio. The operation of a pilot plant of Gulf Power Co. in Florida which will commence shortly using the Chiyoda process should be watched for evaluation of the applicability to flue gas from coal-fired boilers.

The sodium-limestone processes are rather complicated including the decomposition step of sodium sulfate. The processes would need a large land space and relatively high investment cost. On the other hand, sodium consumption is low—about one-fifth of that required for a plant the size of the Wellman-Lord process plant which does not use an oxygen inhibitor. If further crystal growth of calcium sulfite is attained, the process may be useful in the United States to by-produce the sulfite whether it is used for some purpose or discarded.

Both the Kurabo and Dow processes use absorbents with an intermediate pH (3-4) and a moderate L/G ratio (20-40). Operation of commercial plants is about to start for both processes. Although both seem interesting, a better evaluation can be made when the new plants have produced some operating data.

Nippon Kokan (NKK) uses an ammonium sulfite solution at pH 6.8 as the absorbent. Therefore, NKK uses a much smaller L/G ratio than does Kurabo, but it is not easy to eliminate plume, which has been a problem common to ammonia scrubbing processes. The Kurabo process might be advantageous where there are strict regulations on plume. Both NKK and Kurabo processes may be useful also for the production of ammonium sulfate where there is a demand for it.

Other processes Among other processes, the Wellman-Lord process has been most widely used except for the sodium scrubbing processes to by-produce sodium salts. The process is reliable and is easy to control. A main drawback of the process is the high consumption of sodium which necessitates the emission of a considerable amount of wastewater. Further improvement may be desirable before application in the U.S.A. where the restrictions on wastewater are much more severe than in Japan.

It is questionable which is better for sulfuric acid production, the Wellman-Lord process or the magnesia scrubbing process. The latter would give much less wastewater but needs a larger amount of energy for the decomposition of magnesium sulfite and sulfate. Further improvements are desired also in magnesia scrubbing for better operation and less consumption of energy.

Sulfur by-producing processes seem costly except for refineries which have large Claus furnaces to which the recovered SO_2 can be charged. If an oversupply of other desulfurization by-products causes problems, the sulfur by-producing process may become more popular.

References

- 1) Jumpei Ando, Proceedings of Flue Gas Desulfurization Symposium (1973), pp 69-102 and 875-890
- 2) Jumpei Ando, Utilizing and Disposing of Sulfur Products from Flue Gas Desulfurization Processes in Japan, EPA FGD Symposium (Nov. 1974)
- 3) C.P. Quigley, Proceedings of Flue Gas Desulfurization Symposium (1973), pp 605-618
- 4) Masao Endo, Sekko to Sekkai, No. 131, 168-172 (1974)
- 5) Y. Yamamichi and J. Nagao, Sekko to Sekkai, No. 130, 125-129 (1974)

COST COMPARISONS OF FLUE GAS DESULFURIZATION SYSTEMS

G. G. McGlamery and R. L. Torstrick
Tennessee Valley Authority
Muscle Shoals, Alabama

Prepared for Presentation at
Flue Gas Desulfurization Symposium
Sponsored by the Environmental Protection Agency
Atlanta, Georgia
November 4-7, 1974

COST COMPARISONS OF FLUE GAS DESULFURIZATION SYSTEMS¹

G. G. McGlamery and R. L. Torstrick
Tennessee Valley Authority
Muscle Shoals, Alabama

ABSTRACT

This paper describes the results of an EPA-sponsored cost appraisal of the five most advanced flue gas desulfurization processes. Using data available in late 1973 from demonstration-scale systems in progress and a representative set of design and economic premises, detailed, highly visible capital investments and operating costs were estimated for the limestone slurry scrubbing, lime slurry scrubbing, magnesia slurry scrubbing - regeneration to acid, sodium solution scrubbing - regeneration to sulfur, and catalytic oxidation processes.

Itemized capital costs are presented for a base case system (new 500-MW unit burning coal with 3.5% sulfur, 30-year life - 127,500 hours operation, 90% SO₂ removal, fly ash removal included) for each process assuming proven technology, a 3-year construction schedule (1972-1975, midpoint of activity - 1974), no overtime, and an experienced design and construction team. Energy costs reflecting recent worldwide escalation are applied to the operating costs.

In addition to the detailed base cases, numerous supplementary estimates are given to cover such variables as unit size, fuel type, sulfur content of fuel, plant status (new vs. existing), onsite versus offsite solids disposal and SO₂ removal. The five processes are ranked according to the results; the accuracy of the estimates is also discussed.

Using results from recent in-house evaluation studies, additional less-detailed comparisons are presented for limestone scrubbing with benzoic acid, two lime scrubbing options, sodium scrubbing with H₂SO₄ production, and two double-alkali processes.

¹ To be presented at the Flue Gas Desulfurization Symposium, Atlanta, Georgia, November 4-7, 1974.

COST COMPARISONS OF FLUE GAS DESULFURIZATION SYSTEMS

After several years of intensive process development, several power plant stack gas SO₂ removal systems are now advancing from the pilot-plant stage to demonstration-scale applications. With the installation of these larger demonstration-size facilities on power units around the United States, it should now be possible to more accurately predict their costs so that utility executives can better choose between alternatives.

Since the Air Quality Act of 1967 and later the Clean Air Act of 1970, the Federal Environmental Protection Agency has funded research and development on SO₂ removal processes including several conceptual design and cost studies. In these earlier efforts, many design assumptions were necessary and cost estimate accuracy was questionable since technology was in a state of "flux" and available design data were limited. Equipment costs were sketchy since most vendors had yet to fabricate and erect the large gas scrubbing devices required for full-scale systems. Furthermore, very little corrosion data were available to predict materials of construction for the services involved. In many cases, optimism of the process developers tended to maximize process potential and to minimize problem areas such as erosion, scaling, solids disposal, sulfite oxidation, mist elimination, gas reheat, operational turndown, and pH control.

Finally, after many pilot-plant tests and both encouraging and disappointing experiences, five processes have emerged from the many proposed as the initial systems for installation. These are the limestone slurry process and the lime slurry process, both of which are throwaway systems (no salable byproduct), and the magnesia slurry - regeneration to sulfuric acid, sodium solution scrubbing - SO₂ reduction to sulfur, and catalytic oxidation - 80% sulfuric acid processes. In cooperation with participating utilities and process developers, EPA is currently funding large-scale test and demonstration projects on each of these processes. The processes and the associated projects are shown in Table 1.

Now that many of the unknowns for these systems have surfaced, remedies been prescribed, and large-scale projects started, more accurate assessment of process costs should be possible. The objective of an EPA-TVA study completed in early 1974 was to prepare a set of highly visible, detailed capital and operating cost estimates for comparison of the five leading processes on a common, uniform basis. Unfortunately, the rapid construction cost escalation experienced during recent months has once again made it difficult to accurately project total costs; however, comparative costs should be relatively assessable.

Table 1. EPA-SPONSORED STACK GAS DESULFURIZATION DEMONSTRATION SYSTEMS

EPA-sponsored process (byproduct)	Cooperating utility	Process developer	Location	Unit size and type	Expected startup
Limestone slurry scrubbing (sludge)	TVA	Bechtel and others	Shawnee unit 10 Paducah, Ky.	10 MW coal	Under way
Lime slurry scrubbing (sludge)	TVA	Chemico, Bechtel, and others	Shawnee unit 10 Paducah, Ky.	10 MW coal	Under way
Magnesia slurry scrubbing - regeneration (98% sulfuric acid)	Boston Edison	Chemico-Basic	Mystic Station 6 Boston, Mass.	150 MW oil	Completed
Catalytic oxidation (80% sulfuric acid)	Illinois Power	Monsanto	Wood River Station 4 East Alton, Ill.	110 MW coal	Mid-1974
Sodium scrubbing - regeneration (sulfur)	Northern Indiana Public Service Co.	Davy Powergas Allied Chemical	D. H. Mitchell Station 11 Gery, Ind.	115 MW coal	Late 1975

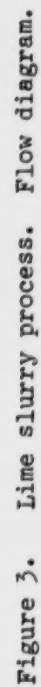
PROCESS DEFINITION

A brief description of these five processes and the organizations supplying representative system data are given below. The process data represent the state of technology in late 1973.

1. Limestone Slurry Scrubbing. Stack gas is washed with a recirculating slurry (pH of 5.8-6.4) of limestone and reacted calcium salts in water using a two-stage (venturi and mobile bed) scrubber system for particulate and SO_2 removal. Limestone feed is wet ground prior to addition to the scrubber effluent hold tank. Calcium sulfite and sulfate salts are withdrawn to a disposal area for discard. Reheat of stack gas to 175°F is provided. Design is based on data taken from EPA-TVA-Bechtel Shawnee test program. A flow diagram and layout drawings are shown in Figures 1 and 2.
2. Lime Slurry Scrubbing. Stack gas is washed with a recirculating slurry (pH of 6.0-8.0) of calcined limestone (lime) and reacted calcium salts in water using two stages of venturi scrubbing. Lime is purchased from "across the fence" calcination operation, slaked, and added to both circulation streams. Calcium sulfite and sulfate is withdrawn to a disposal area for discard. Reheat of stack gas to 175°F is provided. Design is based on data provided by Chemical Construction Corporation (Chemico). The representative flow diagram is shown in Figure 3 and layouts are shown in Figure 4.
3. Magnesia Slurry Scrubbing - Regeneration to H_2SO_4 . Stack gas is washed using two separate stages of venturi scrubbing--the first utilizing water for removal of particulates, and the second utilizing a recirculating slurry (pH 7.5-8.5) of magnesia (MgO) and reacted magnesium - sulfur salts in water for removal of SO_2 . Makeup magnesia is slaked and added to cover only handling losses since sulfates formed are reduced during regeneration. Slurry from the SO_2 scrubber is dewatered, dried, calcined, and recycled during which concentrated SO_2 is evolved to a contact sulfuric acid plant producing 98% acid. Design is based on data supplied by Chemico-Basic Corporation. Shown in Figures 5 and 6 are a flow diagram and layouts of the scrubbing area.
4. Sodium Solution Scrubbing - SO_2 Regeneration and Reduction to Sulfur. Stack gas is washed with water in a venturi scrubber for removal of particulates and then washed in a valve tray scrubber with a recirculating solution of sodium salts in water for SO_2 removal. Makeup sodium carbonate is added to cover losses due to handling and oxidation of sodium sulfite to sulfate. Sodium sulfate crystals are purged from the system, dried, and sold. Water is evaporated from the scrubbing solution using a single-effect evaporator to crystallize and thermally decompose sodium bisulfite, driving off concentrated SO_2 . The resulting sodium sulfite is recycled to the scrubber and the SO_2 is reacted with methane for reduction to elemental sulfur. The regeneration and reduction areas are designed



Figure 1. Limestone slurry process. Flow diagram.



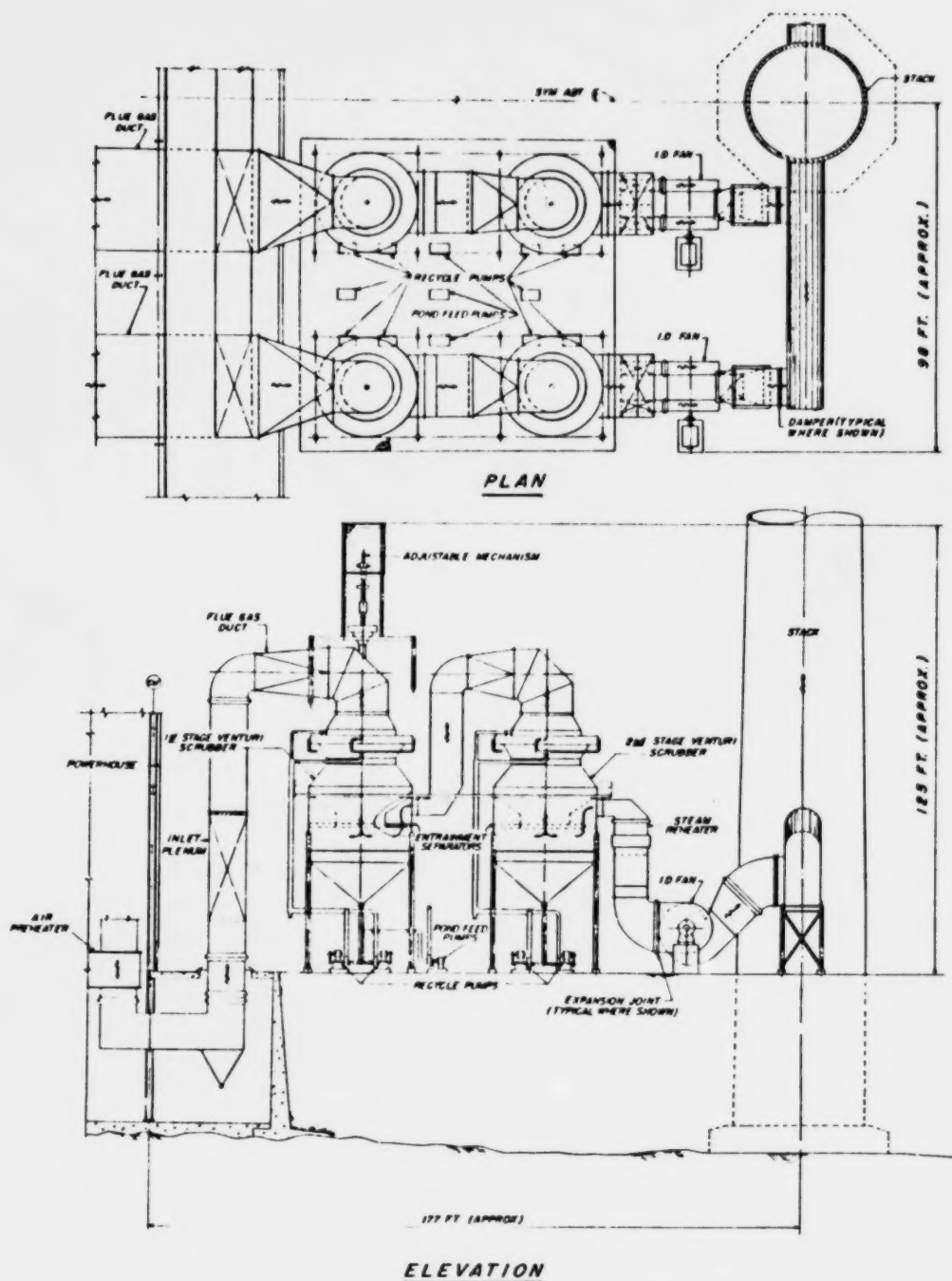


Figure 4. Lime slurry process. Two-stage venturi scrubber system - plan and elevation - base case.

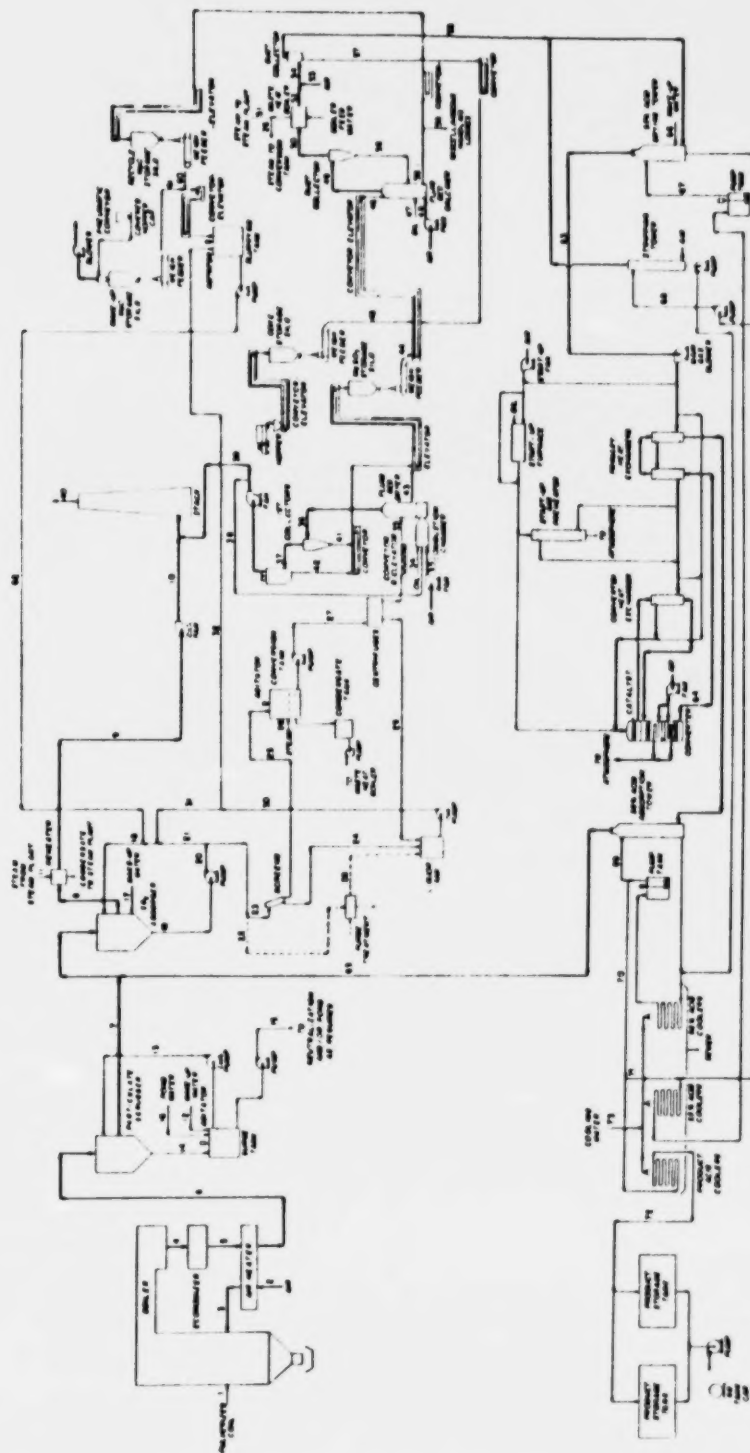


Figure 5. Magnesia slurry - regeneration process. Flow diagram.

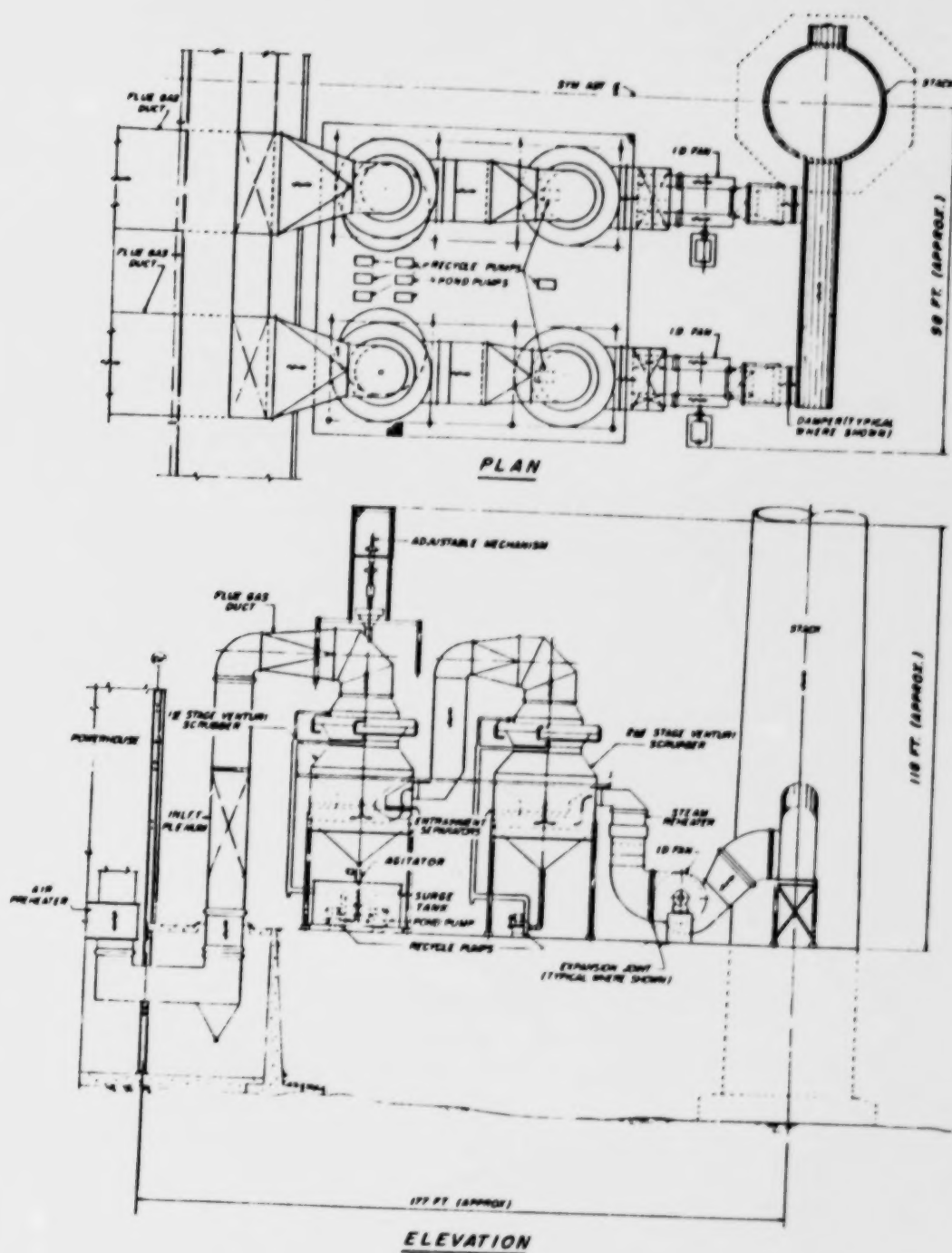


Figure 6. Magnesia slurry - regeneration process. Two-stage venturi scrubber system - plan and elevation - base case.

for 100% of power unit load. Design for the scrubbing - evaporator - crystallizer system is provided by Davy Powergas Inc. (Wellman-Lord process) and data for the SO₂ reduction unit are provided by Allied Chemical Corporation. Figures 7 and 8 describe the processing steps and scrubbing area layout.

5. Catalytic Oxidation. Stack gas is first cleaned of particulates by a high-temperature electrostatic precipitator; then, the SO₂ is catalytically converted to SO₃ and available excess heat is recovered. The SO₃ reacts with moisture in the stack gas to form H₂SO₄ mist which is scrubbed in a packed tower using a recirculating acid stream to yield 80% acid. Mist is removed by a Brink mist eliminator and the clean 254°F gas is exhausted to the stack. Design is based on data supplied by Monsanto Company, developers of Cat-Ox process. Shown in Figures 9-11 are a flowsheet and the process equipment arrangements for a new unit installation.

In the EPA-TVA report, representative flow diagrams, material balances, control diagrams, plant layouts, and equipment arrangements are included for the base case (new 500-MW coal-fired unit, 3.5% S in fuel, 90% SO₂ removal) of each process. Together with detailed equipment descriptions, this background defines the systems estimated.

MAJOR COST FACTORS

From previous economic studies, the following factors are considered to be the most important:

1. Project schedule and location. Project assumed to start in mid-1972 with 3-year construction period ending mid-1975. Midpoint of construction costs mid-1974; Chemical Engineering Cost Index - 160.2. Startup - mid-1975. A midwestern plant location is assumed.
2. Power unit size. Costs for three unit sizes--200, 500, and 1000 MW--are projected.
3. Fuel type. Systems for both coal- and oil-fired units are costed--coal 12,000 Btu/lb, 12% ash, oil - 18,500 Btu/lb, 0.1% ash.
4. Sulfur content of fuel. Costs for three sulfur levels are evaluated for each fuel--2.0, 3.5, and 5.0% for coal; 1.0, 2.5, and 4.0% for oil.
5. Plant status. Although systems for both new and existing power units are evaluated, only a simple, moderately difficult (scrubbing system installed on vacant space beyond the stack) retrofit is estimated since such systems can vary over such a wide range of configurations and restrictions. New units are designed for a 30-year life, 127,500 hours of operation. Costs for new and existing systems are not directly comparable.

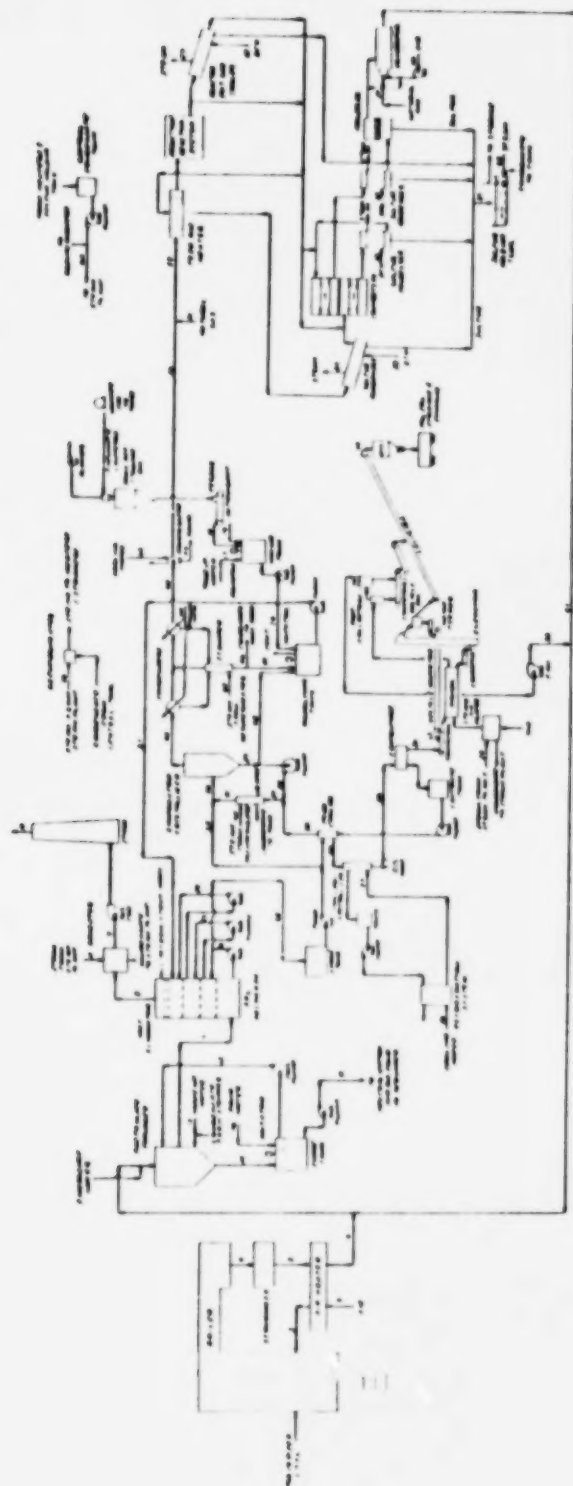


Figure 7. Sodium solution - SO₂ reduction process. Flow diagram.

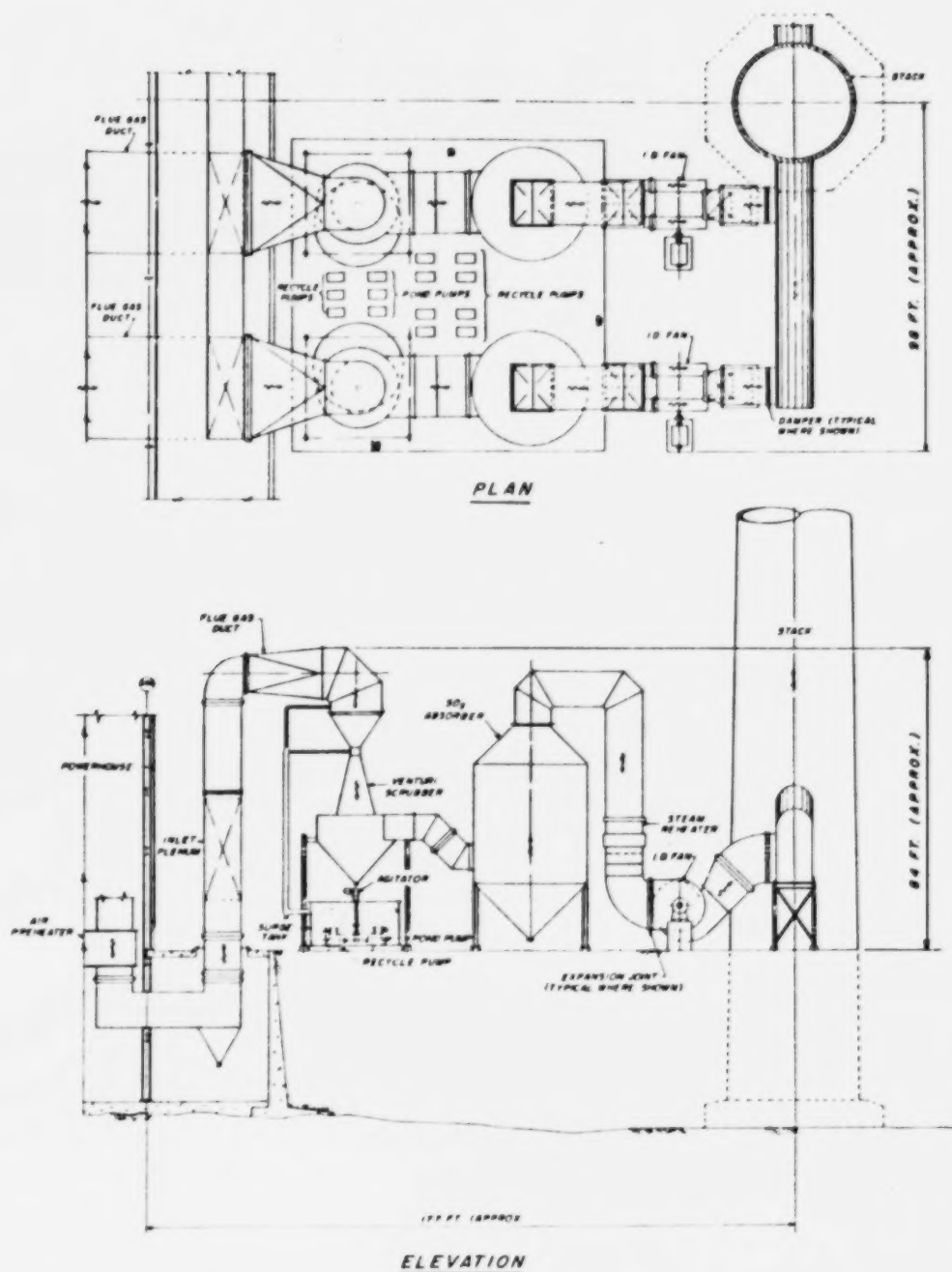
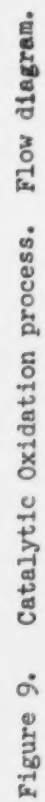


Figure 3. Sodium solution - SO_2 reduction process. Venturi and valve-tray scrubber system - plan and elevation - base case.



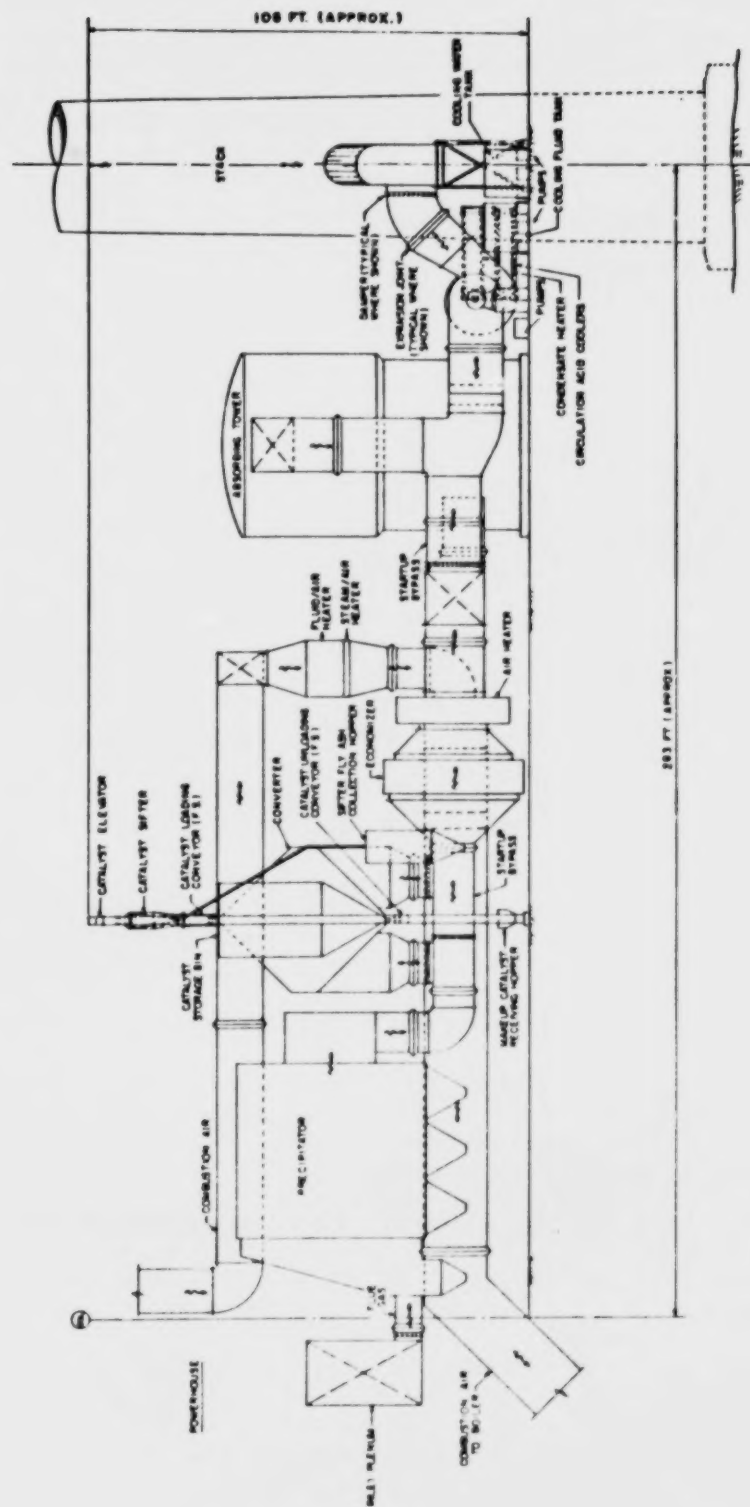


Figure 11. Catalytic oxidation process. SO_2 conversion and absorption system layout - elevation - base case.

6. SO₂ removal. Since all five processes are capable of 90% SO₂ removal and future demands for emission control may exceed present standards, 90% removal is specified as the base value. For those processes in which cost effective design changes could be identified, 80% removal is also projected.
7. Particulate removal. Costs are included for 98.7% particulate removal (to meet EPA standard of 0.1 lb/million Btu heat input) on new coal-fired systems except Cat-Ox which requires 99.9% removal for process reasons (restricted to 0.005 gr/scf prior to entering converter). Existing units are assumed to be already equipped with 98.7% electrostatic precipitators. Because of this provision, the investment and operating cost results for existing coal-fired systems appear lower than for new units. To cover this disparity, a special case is examined where full particulate removal must be added to an existing unit. Oil-fired units do not require dust removal facilities. Existing Cat-Ox installations require incremental additional precipitator to reduce loading to 0.005 gr/scf.
8. Raw materials and catalysts. Assuming startup in 1975, mid-western 1975 delivered prices are projected and sensitivity analysis is used to evaluate variance.
9. Labor. 1975 midwestern operating labor rates are projected. Sensitivity analysis is used to evaluate overall operating labor cost variance. Operating labor is escalated over the life of the project for only a few special cases to show effect.
10. Utilities. Recent energy cost escalation is recognized and 1975 values are projected. Values used in operating cost estimates for utilities supplied by power plant cover all costs for generation including return on investment, depreciation, and income taxes.
11. Maintenance. Various levels are analyzed by sensitivity analysis.
12. Capital charges. Regulated (profit and taxes included) economic basis is used. Annual operating cost estimates utilize a base value of 14.9% of fixed investment (10% cost of money). Level is varied by sensitivity analysis.
13. On-stream time. Annual operating costs are projected for operating times of 7000, 5000, 3500, and 1500 hours per year. Later, these values are used to project a lifetime cost over a predefined 30-year declining operating schedule.
14. Solids disposal. Both onsite ponding and offsite disposal costs are evaluated for limestone and lime processes. Onsite ponding includes prorated costs for calcium solids to cover pumping and piping to and from the pond, plus a 40-foot-deep, clay-lined

pond sized to meet requirements over the remaining life of the power unit. Offsite disposal includes proration of a thickener, filter, piping, pumps, cake conveyors, and loader at the power unit. Offsite charges are levied on a fee per ton of wet solids basis assumed to cover all contractor expenses for hauling, treatment, and final disposal. Sensitivity analysis is used to evaluate variance.

15. Net sales revenue. Base values--\$8/ton 100% H_2SO_4 as 98% H_2SO_4 , \$6/ton 100% H_2SO_4 as 80% H_2SO_4 , \$25/short ton for sulfur, \$20/ton for sodium sulfate are used. Variances are covered by sensitivity analysis.

Other important design and cost assumptions defined for consistent evaluation are:

1. Fly ash disposal facilities proration (ponds, pipes, pumps) not included. Water balance is based on closed-loop operation.
2. System design assumed not "first of kind;" no redundancy is included; only pumps are spared; experienced design and construction team is assumed utilized.
3. Stack gas reheated to 175°F except Cat-Ox process.
4. Product storage - 30 days except for Na_2SO_4 - 7 days.
5. Equipment, material, and construction labor shortages with accompanying overtime pay incentive not considered.

PRESENTATION OF RESULTS

Capital Investment

In the detailed cost study, several methods are used to present the results. For investment data, the following types of displays are presented for each process:

Equipment list, cost, size-scale factor, and data source--presented for base case (new 500-MW coal-fired unit burning coal with 3.5% S). See Table 2 for an example.

Summarized process area equipment and installation cost breakdown--presented for base case and existing 500-MW limestone system. See Table 3 and 4 for examples.

Case variations--area investment summaries presented for each of the 16 case variations. See Tables 5-10 for examples.

Shown in Table 11 is a summary of total investment for the 16 case variations of the five leading processes. The relative ranking of the processes for new coal-fired units (3.5% S) in order of increasing investment is as follows:

Table 2. LIMESTONE SLURRY PROCESS
BASE CASE

Area 1 Material Handling									
Item	No	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1. Unloading hopper No. 1	1	Capacity 94 ft ³ ; 8'- 4" side, 2'-4" bot- tom, 3' deep, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	1,700	Catalytic Inc.	1973	2,000	2,000
2 Limestone feeder No. 1 (vibrating)	1	210 tons/hr, 42" wide x 5' long pan, 2½ hp vibrator included, carbon steel	0.58	Chem. Engr. 3-24-69 Guthrie	2,691	Richardson Engr. Services	1971	3,300	3,300
3 Conveyor (belt) No. 1	1	210 tons/hr, 250 ft/ min, 24" belt, 10' long, 2½ hp motor included, carbon steel	0.81	Fund. of Cost Engr. 1964	1,887	TVA work order D05C30	1964	2,900	2,900
			0.65	Chem. Engr. 3-24-69 Guthrie					
4 Conveyor (belt) No. 2	1	210 tons/hr, 250 ft/ min, 24" belt, 172' long, 20 hp motor included, carbon steel	0.81	Fund. of Cost Engr. 1964	8,948	TVA work order D05P353	1962	13,700	13,700
			0.65	Chem. Engr. 3-24-69 Guthrie					
5 Hoppers under pile	3	7'-4" top, 1'-4" bottom, 3' deep, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	1,300	Chem. Engr. - 3-24-69, Guthrie	1969	1,700	5,100
6 Limestone feeder No. 2 (vibrating)	3	100 tons/hr, 18" wide x 3½' long pan, 1 hp vibrator included, carbon steel	0.58	Chem. Engr. 3-24-69 Guthrie	1,550	Richardson Engr. Services	1973	1,700	5,100
7 Conveyor (belt) No. 3	1	250 ft/min, 18" belt, 135' long, 3 hp motor included, carbon steel	0.65	Chem. Engr. 3-24-69 Guthrie	11,250	Chem. Engr. - 3-24-69, Guthrie	1969	14,800	14,800
			0.81	Fund. of Cost Engr. 1964					
8 Tunnel sump pump	2	5 gpm, 10' head, ¼ hp motor included, carbon steel, neoprene lining		Depends on gpm and head requirements resulting in changes of motor and impeller sizes	560	Catalytic Inc.	1973	600	1,200
9 Elevator No. 1	1	100 tons/hr @ 85 lb/ ft ³ , 16" x 8" x 8½" bucket, 235 ft/min, 15 hp motor included, carbon steel	0.83	Chem. Engr. 3-24-69 Guthrie	8,765	TVA work order D05C30	1964	13,300	13,300
10 Bin	1	5,000 ft ³ , 3/8" carbon steel plate, plus structural steel	0.68	Chem. Engr. 3-24-69 Guthrie	12,650	TVA work order D05C30	1964	19,200	19,200
11 Car shaker	1	Railroad trackside vibrator			4,866	TVA work order D05C30	1965	6,600	6,600
12 Dust collecting system No. 1	1	2,000 cfm inertial separator, XQ cyclone, 2 dust hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69 Guthrie	2,392	Richardson Engr. Services	1973	2,600	2,600
13 Dust collecting system No. 2	1	6,000 cfm inertial separator, XQ cyclone, 2 dust hoppers, fan, and drive	0.80	Chem. Engr. 3-24-69 Guthrie	4,724	Richardson Engr. Services	1973	5,100	5,100
14 Bag filter system	1	14,000 cfm, automatic fabric dust collectors, bag support, shaker sys- tem, isolation damper, external shaker motor and drive, dust hopper, fan and motor for bag filter system (½ in feed preparation area)	0.68	Chem. Engr. 3-24-69 Guthrie	7,926	Richardson Engr. Services	1973	8,500	8,500
Subtotal								103,400	

Table 2. LIMESTONE SLURRY PROCESS (Contd)
BASE CASE

Area 2 Feed Preparation									
Item	No.	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1 Bin discharge feeder	2	12½ tons/hr, 10 wide x 2½ long pan, vibrator included, carbon steel	0.58	Chem Engr 3-24-69 Guthrie	500	Richardson Engr. Services	1973	600	1,200
2 Weigh feeder	2	12½ tons/hr, 18" belt, 14' long, 1½ hp motor included, carbon steel	0.65	Chem Engr 3-24-69 Guthrie	7,216	Catalytic Inc.	1972	8,200	16,400
3 Gyratory crusher	2	12½ tons/hr, 0 x 1½ to ¾, 25 hp motor included	1.20	Chem Engr 3-24-69 Guthrie	10,455	Denver Equip Co	1973	11,300	22,600
4 Elevator No. 2	2	12½ tons/hr, 85 lb/ ft³, 235 ft/min, 24' ctr, 6" x 4" x 4" bucket, 1 hp motor included	0.65	Chem Engr 3-24-69 Guthrie	1,760	Chem Engr 3-24-69, Guthrie	1969	2,400	4,800
5 Wet ball mill	1	300 tons/day, 8' dia x 12' long, from ½" to 200 mesh	0.65	Chem Engr 3-24-69 Guthrie	99,245	Denver Equip Co	1973	108,150	216,300
	2	450 hp motors for ball mill	1.07	Fund of Cost Engr 1964	13,100	Westinghouse	1973	14,200	28,400
6 Mills product tank	1	1,920 gal, 8' diam- eter x 5' high, vertical with open top, carbon steel	0.52	Fund of Cost Engr 1964	2,120	GATX	1971	2,500	2,500
		Lining Neoprene lining for mill product tank			1,400	GATX	1971	800	800
7 Agitator, mills product tank	1	1 hp, neoprene coated	0.50	Chem Engr 3-24-69 Guthrie	1,026	Mixing Equip Co., Inc.	1971	1,200	1,200
			0.26	Fund of Cost Engr 1964					
8 Pumps, mills product tank	2	96 gpm, 58' head, centrifugal, with variable speed drive and 3 hp motor, carbon steel, neoprene lining	Depends on gpm and head requirements resulting in changes of motor and impeller sizes		2,060	Denver Equip Co	1973	2,200	4,400
9 Slurry feed tank	1	46,080 gal, 6,160 ft³, 17' 4" diameter x 27' high, vertical with open top, (4) 1' 5" wide baffles, carbon steel	0.68	Chem Engr 3-24-69 Guthrie	16,000	GATX	1971	19,200	19,200
		Lining ¾" neoprene lining			14,000	GATX	1971	17,000	17,000
10 Agitator, slurry feed tank	1	10 hp, neoprene coated	0.50	Chem Engr 3-24-69 Guthrie	4,717	Mixing Equip Co., Inc.	1971	5,600	5,600
			0.46	Fund of Cost Engr 1964					
11 Pumps, slurry feed tank	2	96 gpm, 58' head, centrifugal, with variable speed drive and 3 hp motor, carbon steel, neoprene lined	Depends on gpm and head requirements resulting in changes of motor and impeller size		2,060	Denver Equip Co	1973	2,200	4,400
12 Dust collecting system	1	8,000 cfm, inertial separator, XQ cyclone, 2 dust hoppers, fan and drive	0.80	Chem Engr 3-24-69 Guthrie	5,894	Richardson Engr Services	1973	6,400	6,400
13 Hoist	1	5 ton electric	0.81	Popper, H.	17,570	Richardson Engr Services	1973	19,000	19,000
14 Bag filter system	1	14,000 cfm, automatic fabric dust collectors, bag support, shaker system, isolation damper, external shaker motor and drive, dust hopper, motor and fan for bag filter system (½ in. materials handling area)	0.68	Chem Engr 3-24-69 Guthrie	7,926	Richardson Engr Services	1973	8,500	8,500
Subtotal								378,700	

Table 2. LIMESTONE SLURRY PROCESS (Contd)
BASE CASE

Area 3 Particulate Scrubbing

Item	No	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1. Tank, particulate scrubber, effluent hold	4	25,700 gal. 3.435 ft ³ , 13' diameter x 26' high, open top, (4) 1' wide baffles, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	13,500	GATX	1971	16,300	65,200
Lining		1/2" neoprene lining			10,000	GATX	1971	12,000	48,000
2 Agitator, effluent hold tank	4	5 hp, neoprene coated	0.26	Fund. of Cost Engr. 1964	3,400	Mixing Equip. Co.	1971	4,000	16,000
3. Pumps, recycle slurry	6	4,900 gpm, 144' head, centrifugal, belt drive, 300 hp motor included, carbon steel, neoprene lining	0.50	Chem. Engr. 3-24-69 Guthrie	13,440	Denver Equip. Co.	1973	14,500	87,000
4. Venturi scrubber	4	With variable throat, 36' long x 5' wide x 20' high, conveying section & throat carpenter 20, remainder 1/2" carbon steel with neoprene lining	0.60	Universal Oil Products	133,000	Universal Oil Products	1971	150,000	600,000
5 Venturi & MRA sump	4	28' long x 41' wide x 13' high, 1/2" carbon steel, neoprene lining (1/2 in SO ₂ scrubbing area)	0.68	Chem. Engr. 3-24-69 Guthrie	49,500	Universal Oil Products	1971	57,500	230,000
6 Soot blowers	20		1.00	TVA	3,500	Widows Creek - TVA	1971	4,000	80,000
Subtotal									1,126,200

Table 2. LIMESTONE SLURRY PROCESS (Contd)
BASE CASE

Area 4. SO ₂ Scrubbing									
Item	No	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1. ICA scrubber	4	SO ₂ absorber, mobile bed, with demister, 41' long x 16' wide x 41' high, 1/2" carbon steel with neoprene lining, 316 S.S. grids, high density poly. spheres, 1 RP spray headers, chevron vane mist eliminators	0.80	Universal Oil Products	227,000	Universal Oil Products	1971	300,000	1,200,000
2. Venturi & MBA sump	4	28' long x 41' wide x 13' high, 1/2" carbon steel, neoprene lining (1/2 in particulate scrubbing area)	0.68	Chem. Engr. 3-24-69 Guthrie	49,500	Universal Oil Products	1971	57,500	230,000
3. Tank, absorber effluent hold	4	240,000 gal, 32,088 ft ³ 40' diameter x 26' high, open top, field erected, carbon steel	0.68	Chem. Engr. 3-24-69 Guthrie	38,327	GATX	1972	45,100	180,400
Lining		1/4" neoprene lining			32,000	GATX	1972	38,600	154,400
4. Agitator, SO ₂ absorber hold tank	4	50 hp, neoprene coated	0.50	Chem. Engr. 3-24-69 Guthrie	12,245	Catalytic Inc.	1973	13,250	53,000
5. Pumps, SO ₂ absorber recycle slurry	10	11,500 gpm, 105' head, centrifugal, neoprene lined, belt drive, 500 hp motor included		Depends on gpm and head requirements resulting in changes of motor and impeller size	23,000	Allen, Sherman, Hoff	1971	27,100	271,000
6. Pumps, makeup water	2	1,240 gpm, 150' head, vertical multistage turbine, 100 hp motor included		Depends on gpm and head requirements resulting in changes of motor and impeller size	2,369	Richardson Engr. Services	1973	2,600	5,200
7. Soot blowers	40		1.00	TVA	3,500	Widows Creek TVA	1971	4,000	160,000
Subtotal									2,254,000

Table 2. LIMESTONE SLURRY PROCESS (Contd)
BASE CASE

Area 5 Reheat

Item	No	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1 Gas reheater	4	Tube type, 2.028 ft 13,900 lb. $\frac{1}{2}$ of tubes made of inconel 625 and remaining $\frac{1}{2}$ made of cor-ten	0.80	Chem Engr Guthrie	70,000	Widows Creek TVA	1971	82,500	330,000
2 Soot blowers	20		1.00	TVA	3,500	Widows Creek TVA	1971	4,000	80,000
Subtotal									410,000

Area 6 Gas Handling

Item	No	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1 Fan	4	41" 1,200 rpm, 2,905 bhp, 3,250 hp motor included with insulation (Cost is difference between 41" and 15" fan Remainder is allocated to boiler)	0.68	Chem Engr Guthrie	59,710	Widows Creek TVA	1971	71,150	284,600
Subtotal									284,600

Table 2. LIMESTONE SLURRY PROCESS (Contd)
BASE CASE

Area 7 Solids Disposal									
Item	No	Description	Size-cost scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1 Pond feed tank	1	63,000 gal, field erected, 8,423 ft ³ , 21' diameter x 26' high, vertical with open top, carbon steel, 1' - 8" x 26' baffles	0.68	Chem. Engr 3-24-69 Guthrie	20,000	GATX	1971	24,000	24,000
Lining		¾" neoprene lining			16,120	GATX	1971	19,000	19,000
2 Agitator	1	7½ hp, neoprene coated	0.50 0.47	Chem. Engr 3-24-69 Fund of Cost Engr 1964	4,717	Mixing Equip Co., Inc.	1971	5,400	5,400
3 Pumps, pond feed tank ^a	4	1,128 gpm @ 75' head, neoprene lined with 50 hp motor		Depends on gpm and head requirements resulting in changes of motor and impeller size	1,850	Denver Equip Co.	1973	2,225	8,900
4 Pumps, recycle pond water ^a	2	1,000 gpm @ 150' head, multistage turbine, cast iron bowl, stainless steel impellers, with 75 hp motor		Depends on gpm and head requirements resulting in changes of motor and impeller size	1,100	Richardson Engr Services	1973	1,450	2,900
Subtotal									60,200

^aCost for this equipment prorated for fly ash removal

Area 8 - Utilities

Note: There is no process equipment in this area

Table 2. LIMESTONE SLURRY PROCESS (Contd)
BASE CASE

Area 9. Services									
Item	No.	Description	Size or scale factor	Factor source	Base cost	Base cost source	Date of cost	Projected equipment each	1974 cost total
1 Payloader	1	Gasoline type, 2 yd ³			21,000	Chem. Engr. 3 24-69, Guthrie	1969	24,800	24,800
2 Plant vehicles		(allocation)			10,000		1974	10,000	10,000
3 Maint. & instrument shop equipment		Office, machine tools, and machine shop equipment			22,400	Chem. Engr. 3 24-69, Guthrie	1969	26,400	26,400
4 Service building equipment		Equipment for lab, locker room, motor control room, restrooms			29,700	Chem. Engr. 3 24-69, Guthrie	1969	35,000	35,000
5 Stores equipment		Office equipment, shelving, etc.			9,000	Chem. Engr. 3 24-69, Guthrie	1969	10,600	10,600
Subtotal									106,800

Table 3. LIMESTONE SLURRY PROCESS - TOTAL CAPITAL INVESTMENT REQUIREMENTS
BASE CASE^a SUMMARY - PROCESS EQUIPMENT AND INSTALLATION ANALYSIS
(Thousands of Dollars)

Direct Costs										
Equipment	Raw material building	Field production	Particulars working	50% working	Release	Farm diagnosis	Utilities	Services	Construction facilities	Percent of total capital investment
									5%	Total
103	139	1,126	2,254	410	285	60	-	107	-	4,724
26	57	114	335	80	34	7	-	23	-	677
1	14	102	341	11	-	86	6	-	-	581
1	18	97	313	20	-	16	11	-	-	346
15	6	733	377	-	166	-	-	-	-	1,277
6	6	582	418	-	53	-	-	-	-	1,067
14	7	12	24	-	4	0	-	-	-	62
81	38	48	81	-	21	5	-	-	-	274
-	-	-	-	-	-	-	-	-	-	3,387
11	-	64	65	-	-	-	-	-	-	180
10	-	18	41	-	-	2	-	-	-	81
19	58	61	101	1	133	50	10	-	-	433
61	92	73	132	1	155	184	10	-	-	710
8	49	77	142	21	13	11	16	-	-	389
4	25	34	61	11	5	5	8	-	-	196
1	1	3	3	-	-	2	3	-	-	13
5	6	15	13	1	6	13	5	-	-	39
-	65	-	-	-	-	-	-	-	-	170
-	67	-	-	-	-	-	-	-	-	107
6	1	2	2	-	-	199	-	16	-	420
-	-	-	-	-	-	-	-	-	-	765
419	899	3,203	4,565	556	854	3,923	67	618	763	16,089
-	-	-	-	-	-	-	-	-	-	100.0
Indirect Costs										
18	81	288	427	10	77	153	6	57	69	1,446
46	99	352	522	41	94	432	6	70	84	1,748
21	45	160	237	28	43	186	3	32	38	803
42	90	320	474	16	85	382	7	64	77	1,687
546	1,214	6,323	6,403	551	1,153	5,796	91	861	1,033	21,693
45	97	346	513	60	92	424	7	49	82	1,315
45	97	346	513	60	92	424	7	49	82	1,315
-	-	-	-	-	-	-	-	-	-	765
636	1,408	5,015	5,411	671	1,337	6,446	106	999	1,197	25,183
26	56	199	295	3	3	24	0	6	6	48
Percent of total capital investment										

100-MW new coal-fired power unit, 15% off first 100% SO_2 removal, on-site solids disposal. Stack gas return to 175 ft by adiabatic steam reheat. Dripback pond located 1 mile from power plant. Deliveries start location represents project beginning mid-1972, ending mid-1975. Average cost basis for ending mid-1974 minimum in process delivery; only pumps are spared. Investment requirements for disposal of fly ash excluded. Construction labor overages with accompanying overheads not considered.

Table 4. LIMESTONE SLURRY PROCESS - TOTAL CAPITAL INVESTMENT REQUIREMENTS
EXISTING CASE^a SUMMARY - PROCESS EQUIPMENT AND INSTALLATION ANALYSIS
(Thousands of Dollars)

Direct Costs	Fuel handling	SO ₂ scrubbing	Reheat	Solid disposal	Utilities	Services	Construction facilities %	Percent of	
								direct investment	total capital investment
Equipment									
Material	105	190	2,210	186	450	67	31	108	
Labor	37	72	436	23	84	4	57	32	3,847 27.2
Paint and insulation									715 5.2
Material	1	14	353	9	87	12	-	-	476 3.4
Labor	1	56	403	27	99	19	-	-	585 4.1
Directly, charges, and supports									772 5.5
Material	15	6	383	-	188	-	-	-	791 5.6
Labor	10	6	387	-	186	-	-	-	54 0.4
Concrete foundations									307 2.2
Material	14	7	34	-	7	1	-	-	3,202 22.9
Labor	103	48	103	1	39	4	-	-	54 0.4
Excavation, site preparation, rubble, roads, and pond									307 2.2
Structural									1,202 8.6
Material	11	-	86	-	-	-	-	-	97 0.7
Labor	13	-	52	-	-	-	-	-	46 0.3
Electrical	40	59	102	8	131	51	52	-	447 3.2
Material	105	117	142	18	194	233	67	-	876 6.2
Labor									306 2.1
Instruments									183 1.3
Material	8	50	165	15	11	20	-	-	11 0.1
Labor	5	32	103	14	9	6	12	-	58 0.4
Paint and miscellaneous									180 1.3
Material	1	1	3	-	2	4	-	-	135 0.9
Labor									314 2.2
Buildings									672 4.8
Material	-	64	-	-	-	-	-	-	100 0
Labor	-	85	-	-	-	-	-	-	1592 11.3
Land	6	1	2	-	291	-	14	-	672 4.8
Construction facilities									14,116 100.0
Subtotal direct investment	482	1,000	3,243	323	1,710	3,411	335	149	61.1
Indirect Costs									
Engineering design and supervision	49	100	534	32	171	361	34	74	1,412 10.0
Construction field expense	63	130	682	42	222	449	44	96	1,835 13.0
Construction office	34	70	367	22	120	253	23	52	988 7.0
Contingency	23	110	577	16	188	397	37	81	1,553 11.0
Subtotal fixed investment	681	1,410	7,393	455	2,411	5,091	473	1,043	19,304 141.0
Allowance for startup and modifications	34	113	592	36	193	407	18	83	1,592 11.3
Interest during construction	34	113	592	36	193	407	18	83	1,592 11.3
Total capital investment	789	1,636	8,577	527	2,797	5,905	549	1,209	23,088 163.6
Percent of total capital investment	3.4	7.1	37.1	2.3	12.1	25.6	2.4	5.2	100.0

^aBased on 100,000 existing coal-fired power unit, 3.5% S in fuel, 90% SO₂ removal, on-site solid disposal.
Stack gas scrubbed 1.1% S in fuel, 90% SO₂ removal, on-site solid disposal.
Disposal pond located 1 mile from power plant.
Minimum plant location represents project beginning mid-1975. Average cost base for existing, mid-1974.
Minimum in-process storage, only pumps are stored.
Remaining life of power unit, 25 yr.
Investment requirements for removal and disposal of fly ash excluded.
Construction labor shortage with accompanying overtime per incentive not considered.

Table 5. LIMESTONE SLURRY PROCESS
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Investment, \$	Percent of subtotal direct investment
Limestone receiving and storage (hoppers, feeders, conveyors, elevators, and bins)	419,000	2.6
Feed preparation (feeders, crushers, elevators, ball mills, tanks, and pumps)	899,000	5.6
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, and pumps)	3,203,000	19.9
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, effluent hold tanks, agitators, pumps, and exhaust gas ducts to inlet of fan)	4,745,000	29.5
Stack gas reheat (4 indirect steam reheaters)	556,000	3.5
Fans (4 fans including exhaust gas ducts and dampers between fan and stack gas plenum)	854,000	5.3
Calcium solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, pond, liner, and pond water return pumps)	3,923,000	24.4
Utilities (instrument air generation and supply system, plus distribution systems for obtaining process steam, water, and electricity from the power plant)	67,000	0.4
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	638,000	4.0
Construction facilities	765,000	4.8
Subtotal direct investment	16,069,000	100.0
Engineering design and supervision	1,446,000	9.0
Construction field expense	1,768,000	11.0
Contractor fees	803,000	5.0
Contingency	1,607,000	10.0
Subtotal fixed investment	21,693,000	135.0
Allowance for startup and modifications	1,735,000	10.8
Interest during construction (8% annual rate)	1,735,000	10.8
Total capital investment	25,163,000	156.6

^a Basis:

Stack gas reheat to 175°F by indirect steam reheat.
Disposal pond located 1 mile from power plant.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 6. LIME SLURRY PROCESS
VENTURI - VENTURI SCRUBBING SCHEME
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Investment, \$	Percent of subtotal direct investment
Lime receiving and storage (bins, feeders, conveyors, and elevators)	795,000	5.5
Feed preparation (conveyors, slakers, tanks, agitators, and pumps)	387,000	2.7
Particulate - sulfur dioxide scrubbers and inlet ducts (4 scrubbers including common feed plenum and pumps)	4,017,000	28.1
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, pumps, and exhaust gas ducts to inlet of fans)	3,153,000	22.0
Stack gas reheat (4 indirect steam reheaters)	542,000	3.8
Fans (4 fans including exhaust gas ducts and dampers between fan and stack gas plenum)	767,000	5.4
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps, pond, liner, and pond water return pumps)	3,356,000	23.4
Utilities (instrument air generation and supply system, plus distribution systems for obtaining process steam, water, and electricity from the power plant)	67,000	0.5
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	552,000	3.8
Construction facilities	682,000	4.8
Subtotal direct investment	14,318,000	100.0
Engineering design and supervision	1,289,000	9.0
Construction field expense	1,575,000	11.0
Contractor fees	716,000	5.0
Contingency	1,432,000	10.0
Subtotal fixed investment	19,330,000	135.0
Allowance for startup and modifications	1,546,000	10.8
Interest during construction (8%/annum rate)	1,546,000	10.8
Total capital investment	22,422,000	156.6

^a Basis:

- Stack gas reheat to 175°F by indirect steam reheat.
- Disposal pond located 1 mile from power plant.
- Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
- Minimum in process storage; only pumps are spared.
- Investment requirements for disposal of fly ash excluded.
- Construction labor shortages with accompanying overtime pay incentive not considered.

Table 7. LIME SLURRY PROCESS
VENTURI - VENTURI SCRUBBING SCHEME
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW existing coal-fired power unit, 3.5% in fuel;
90% SO₂ removal; onsite solids disposal)

	Investment, \$	Percent of subtotal direct investment
Lime receiving and storage (bins, feeders, conveyors, and elevators)	876,000	5.5
Feed preparation (conveyors, slakers, tanks, agitators, and pumps)	436,000	2.7
First stage sulfur dioxide scrubbers and ducts (4 scrubbers including common feed plenum, pumps, and all ductwork between outlet of supplemental fans and the scrubbers)	4,565,000	28.7
Second stage sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, pumps, and all ductwork between scrubbers and stack gas plenum)	3,797,000	23.9
Stack gas reheat (4 direct oil-fired reheaters)	305,000	1.9
Fans (4 fans including ducts and dampers between tie-in to existing duct and inlet to supplemental fan)	1,143,000	7.2
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps, pond, liner, and pond water return pumps)	3,049,000	19.1
Utilities (instrument air generation and supply system, fuel oil storage and supply system, and distribution systems for obtaining process water and electricity from the power plant)	335,000	2.1
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	649,000	4.1
Construction facilities	758,000	4.8
Subtotal direct investment	15,913,000	100.0
Engineering design and supervision	1,591,000	10.0
Construction field expense	2,069,000	13.0
Contractor fees	1,114,000	7.0
Contingency	1,750,000	11.0
Subtotal fixed investment	22,437,000	141.0
Allowance for startup and modifications	1,795,000	11.3
Interest during construction (8% annum rate)	1,795,000	11.3
Total capital investment	26,027,000	163.6

^a Basis:

Stack gas reheat to 175°F by direct oil-fired reheat.
Disposal pond located 1 mile from power plant.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Remaining life of power unit, 25 yr.
Investment requirements for removal and disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 8. MAGNESIA SLURRY - REGENERATION PROCESS^a
SUMMARY OF ESTIMATED FIXED INVESTMENTS

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; 15.8 tons/hr 100% H₂SO₄)

	Investment, \$	Percent of subtotal direct investment
Magnesium oxide and coke receiving and storage (pneumatic conveyor and blower, hoppers, conveyors, elevators, and storage silos)	192,000	1.2
Feed preparation (weigh feeders, conveyors, elevators, slurrying tank, agitator, and pumps)	238,000	1.5
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, pumps, and fly ash neutralization facilities)	3,966,000	24.3
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, pumps, and exhaust gas ducts to inlet of fan)	2,592,000	15.9
Stack gas reheat (4 indirect steam reheaters)	509,000	3.1
Fans (4 fans including exhaust gas ducts and dampers between fan and stack gas plenum)	741,000	4.5
Slurry processing (screens, tanks, pumps, agitators and heating coils, centrifuges, conveyors, and elevators)	711,000	4.3
Drying (fluid bed dryer, fans, combustion chamber, dust collectors, conveyors, elevators, and MgSO ₃ storage silo)	972,000	5.9
Calcining (fluid bed calciner, fans, weigh feeders, conveyors, elevators, waste heat boiler, dust collectors, and recycle MgO storage silo)	1,108,000	6.8
Sulfuric acid plant (complete contact unit for sulfuric acid production including dry gas purification system)	3,197,000	19.6
Sulfuric acid storage (storage and shipping facilities for 30 days production of H ₂ SO ₄)	278,000	1.7
Utilities (instrument air generation and supply system, fuel oil storage and supply system, and distribution systems for obtaining process steam, water, and electricity from power plant)	269,000	1.6
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	783,000	4.8
Construction facilities	778,000	4.8
Subtotal direct investment	16,334,000	100.0
Engineering design and supervision	1,797,000	11.0
Construction field expense	1,797,000	11.0
Contractor fees	817,000	5.0
Contingency	1,633,000	10.0
Subtotal fixed investment	22,378,000	137.0
Allowance for startup and modifications	2,238,000	13.7
Interest during construction (8%/annum rate)	1,790,000	11.0
Total capital investment	26,406,000	161.7

^a Basis:

Stack gas reheat to 175°F by indirect steam reheat.
Midwest plant location represents projects beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Fly ash slurry neutralized before disposal; closed loop water utilization for first stage.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 9. SODIUM SOLUTION - SO₂ REDUCTION PROCESS^a
SUMMARY OF ESTIMATED FIXED INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; 4.7 tons/hr sulfur)

	Investment, \$	Percent of subtotal direct investment
Soda ash and antioxidant receiving, storage, and preparation (pneumatic conveyor and blower, feeders, mixing tank, agitator, and pumps)	225,000	1.2
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, pumps, and fly ash neutralization facilities)	3,846,000	20.4
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, pumps, and exhaust gas ducts to inlet of fan)	4,269,000	22.6
Stack gas reheat (4 indirect steam reheaters)	539,000	2.9
Fans (4 fans including exhaust gas ducts and dampers between fans and stack gas plenum)	889,000	4.7
Purge treatment (refrigeration system, chiller-crystallizer, feed coolers, centrifuge, rotary dryer, steam/air heater, fan, dust collectors, feeders, tanks, agitators, pumps, conveyors, elevator, and bins)	1,473,000	7.8
Sulfur dioxide regeneration (evaporator-crystallizers, heaters, condensers, strippers, compressors, desuperheater, tanks, agitators, and pumps)	2,717,000	14.4
Sulfur dioxide reduction unit	2,921,000	15.5
Sulfur storage (storage and shipping facilities for 30 days production of molten sulfur)	227,000	1.2
Utilities (instrument air generation and supply system, and distribution systems for obtaining process steam, water, and electricity from power plant)	195,000	1.0
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	662,000	3.5
Construction facilities	898,000	4.8
Subtotal direct investment	18,861,000	100.0
Engineering design and supervision	2,075,000	11.0
Construction field expense	2,075,000	11.0
Contractor fees	943,000	5.0
Contingency	1,886,000	10.0
Subtotal fixed investment	25,840,000	137.0
Allowance for startup and modifications	2,584,000	13.7
Interest during construction (8%/annum rate)	2,067,000	11.0
Total capital investment	30,491,000	161.7

^a Basis:

Stack gas reheat to 175°F by indirect steam reheat.

Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.

Minimum in process storage; only pumps are spared.

Fly ash slurry neutralized before disposal; closed loop water utilization for first stage.

Investment requirements for disposal of fly ash excluded.

Construction labor shortages with accompanying overtime pay incentive not considered.

Table 10. CATALYTIC OXIDATION PROCESS
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; 15.7 tons/hr 100% H₂SO₄)

	Investment, \$	Percent of subtotal direct investment
Converter and absorber startup bypass ducts and dampers	491,000	1.9
Electrostatic precipitators and inlet ducts (4 high temperature electrostatic precipitators including common feed plenum)	8,736,000	34.4
Sulfur dioxide converters and ducts (4 converters including catalyst sifter, hopper, storage bin, conveyors, and elevators)	2,145,000	8.5
Heat recovery and ducts (4 steam/air heaters and 4 fluid/air heaters including ducts between economizers and air heaters, and combustion air ducts and dampers between powerhouse and air heaters; investment credit for use of smaller air heaters included)	1,475,000	5.8
Fans (4 ID fans including exhaust gas ducts and dampers between ID fans and stack gas plenum)	1,412,000	5.6
Sulfuric acid absorbers and coolers (2 absorbers including mist eliminators, coolers, tanks, pumps, and ducts and dampers between air heaters and ID fans)	8,917,000	35.2
Sulfuric acid storage (storage and shipping facilities for 30 days production of H ₂ SO ₄)	409,000	1.6
Utilities (instrument air generation and supply system, and distribution systems for obtaining process steam, water, and electricity from power plant)	57,000	0.2
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	518,000	2.0
Construction facilities	1,208,000	4.8
Subtotal direct investment	25,368,000	100.0
Engineering design and supervision	2,790,000	11.0
Construction field expense	2,790,000	11.0
Contractor fees	1,268,000	5.0
Contingency	2,527,000	10.0
Subtotal fixed investment	34,755,000	137.0
Allowance for startup and modifications	3,475,000	13.7
Interest during construction (8%/annum rate)	2,780,000	11.0
Total capital investment excluding catalyst	41,008,000	161.7
Catalyst	1,728,000	6.8
Total capital investment	42,736,000	168.5

^a Basis:

Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.

Only pumps are spared.

Investment requirements for disposal of fly ash excluded.

Construction labor shortages with accompanying overtime pay incentive not considered.

Table 11. SUMMARY - TOTAL CAPITAL INVESTMENT REQUIREMENTS^a

Case	Years life	Limestone process		Lime process		Magnesia process		Sodium process		Cat-Ox process	
		\$	\$/kW	\$	\$/kW	\$	\$/kW	\$	\$/kW	\$	\$/kW
Coal-fired power unit											
90% SO ₂ removal; onsite solids disposal	30	13,031,000	65.2	11,749,000	58.7	14,139,000	70.7	16,198,000	81.0	19,537,000	97.7
200 MW new, 3.5% sulfur	20	11,344,000	56.7	13,036,000	65.2	14,372,000	71.9	17,149,000	85.7	17,735,000	99.7
200 MW existing, 3.5% sulfur	25	23,083,000	46.2	26,027,000	52.1	26,026,000	52.1	31,208,000	62.4	37,907,000	75.8
500 MW new, 3.0% sulfur	30	22,600,000	45.2	20,232,000	40.5	22,958,000	45.9	26,706,000	53.4	42,520,000	95.0
500 MW new, 3.5% sulfur	30	25,165,000	50.3	22,422,000	44.8	26,406,000	52.8	30,491,000	61.0	42,736,000	95.5
500 MW new, 5.0% sulfur	30	27,343,000	54.7	24,272,000	48.5	29,355,000	58.7	33,709,000	67.4	42,928,000	95.9
1000 MW existing, 3.5% sulfur	25	35,133,000	35.1	38,133,000	38.1	38,717,000	38.7	47,721,000	47.7	62,913,000	62.9
1000 MW new, 3.5% sulfur	30	37,725,000	37.7	32,765,000	32.8	38,865,000	38.9	45,832,000	45.8	69,889,000	69.9
90% SO ₂ removal; onsite solids disposal	30	24,267,000	48.5	21,586,000	43.2	25,568,000	51.1	29,127,000	58.3	-	-
500 MW new, 3.5% sulfur	30	20,532,000	41.1	18,323,000	36.6	-	-	-	-	-	-
90% SO ₂ removal; offsite solids disposal	25	29,996,000	60.0	26,090,000	52.2	32,213,000	64.4	37,997,000	75.9	43,816,000	87.6
500 MW existing, 3.5% sulfur	30	3,263,000	41.3	9,482,000	47.4	8,861,000	44.3	10,324,000	51.6	13,069,000	65.3
200 MW new, 2.5% sulfur	30	12,935,000	25.9	15,961,000	31.9	12,695,000	25.4	15,198,000	30.4	28,067,000	56.1
500 MW new, 1.0% sulfur	30	15,473,000	30.9	18,148,000	36.3	16,080,000	32.2	18,949,000	37.9	28,277,000	56.6
500 MW new, 2.5% sulfur	30	17,431,000	35.0	19,361,000	39.7	18,765,000	37.5	21,893,000	43.8	28,449,000	56.9
500 MW new, 4.0% sulfur	25	19,657,000	37.3	21,317,000	43.6	20,376,000	40.8	24,445,000	48.9	32,824,000	65.6
1000 MW existing, 2.5% sulfur	30	23,584,000	23.4	26,341,000	26.3	23,656,000	23.7	28,765,000	28.8	46,356,000	46.4
1000 MW new, 2.5% sulfur	30	23,584,000	23.4	26,341,000	26.3	23,656,000	23.7	28,765,000	28.8	46,356,000	46.4
Oil-fired power unit											
90% SO ₂ removal; onsite solids disposal	30	3,263,000	41.3	9,482,000	47.4	8,861,000	44.3	10,324,000	51.6	13,069,000	65.3
200 MW new, 2.5% sulfur	30	12,935,000	25.9	15,961,000	31.9	12,695,000	25.4	15,198,000	30.4	28,067,000	56.1
500 MW new, 1.0% sulfur	30	15,473,000	30.9	18,148,000	36.3	16,080,000	32.2	18,949,000	37.9	28,277,000	56.6
500 MW new, 2.5% sulfur	30	17,431,000	35.0	19,361,000	39.7	18,765,000	37.5	21,893,000	43.8	28,449,000	56.9
500 MW new, 4.0% sulfur	25	19,657,000	37.3	21,317,000	43.6	20,376,000	40.8	24,445,000	48.9	32,824,000	65.6
1000 MW existing, 2.5% sulfur	30	23,584,000	23.4	26,341,000	26.3	23,656,000	23.7	28,765,000	28.8	46,356,000	46.4
1000 MW new, 2.5% sulfur	30	23,584,000	23.4	26,341,000	26.3	23,656,000	23.7	28,765,000	28.8	46,356,000	46.4

^a Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974. Minimum in process storage; only pumps are spared. Investment requirements for disposal of fly ash excluded. Construction labor shortages with accompanying overtime pay incentive not considered.

^b All Cat-Ox installations require particulate removal to 0.005 gr/scf prior to entering converter. Because existing units are assumed to already meet EPA standards (0.1 lb particulate/Mcf Btu of heat input). Only incremental additional precipitator is required.

1. Lime slurry
2. Limestone slurry
3. Magnesia slurry - regeneration
4. Sodium solution - SO₂ reduction
5. Catalytic oxidation

Shown in Figures 12 and 13 are the effects of unit size and sulfur content of coal on investment for installations on new units. In comparing these rankings, it should be pointed out that the lime slurry process does not provide facilities for calcining limestone, and includes a minimum of facilities for storage of the calcined material.

An important result derived by comparing the investment projections in this study is the relatively minor investment savings realized (3.2-4.5%) in designing for 80% SO₂ removal as compared to 90%.

The projected investments for limestone and lime slurry processes designed for offsite solids disposal represent savings equivalent to approximately 13% of the projected onsite investment. However, these projections do not include the capital for offsite waste treatment facilities.

Accuracy. In reviewing the capital investment estimates for these five processes, it should be understood that the base case process definitions and estimates represent a generalized cost. As an indication of how the project scope and corresponding investment could vary, the effect of "add ons" to the limestone slurry process base case estimate is shown in Table 12. Such "add ons" as delays in equipment delivery, physical space limitations, additional redundancy requirements, taller stacks, inclusion of a closed-loop fly ash disposal pond, and overtime to accelerate the project completion schedule could more than double the investment for the limestone slurry process. Excluding these additional costs, but considering only the data available for this appraisal and the depth of investigation, the estimated investment accuracy for each of the processes is given in Table 13.

Table 13. PROJECTED INVESTMENT ESTIMATES ACCURACY
BASED ON AVAILABLE DATA AND DEPTH OF INVESTIGATION

Process	% range
Limestone slurry	+20 to -5
Lime slurry	+20 to -10
Magnesia slurry - regeneration	+25 to -15
Sodium solution - regeneration	+25 to -10
Catalytic oxidation (Cat-Ox)	+20 to -10

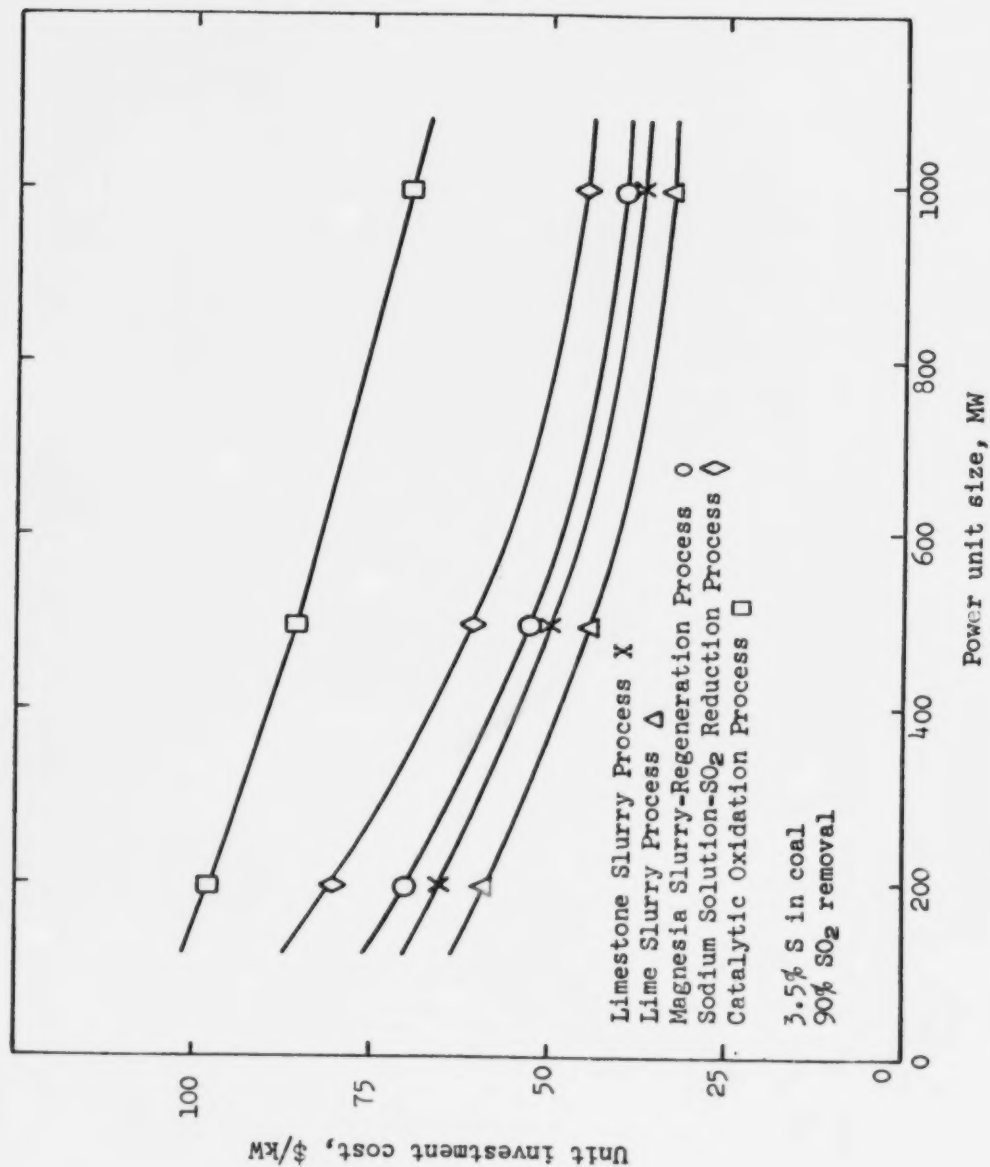


Figure 12. New coal-fired units - the effect of power unit size on unit investment cost, dollars per kilowatt

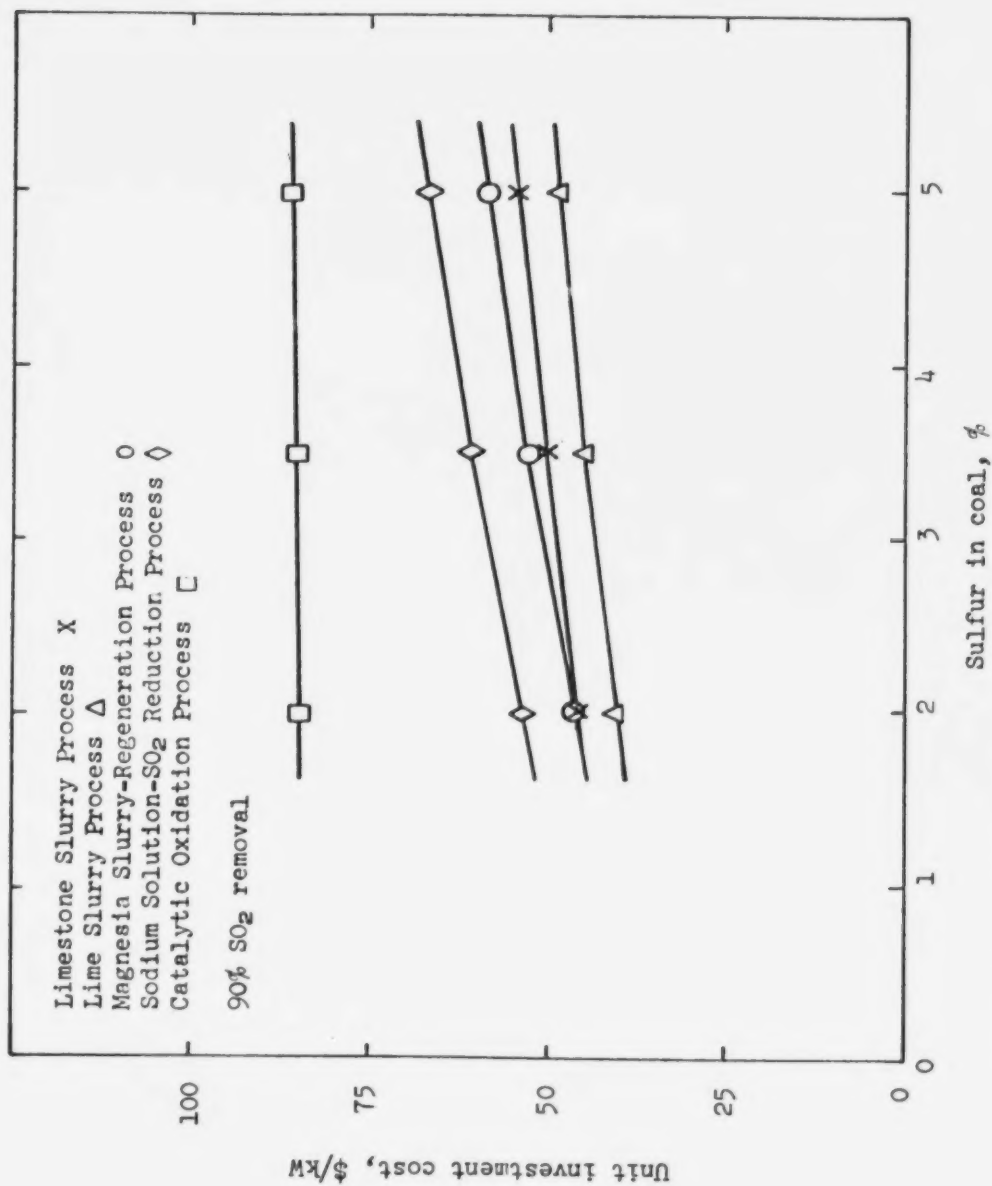


Figure 13. 500-MW new coal-fired units - effect of sulfur content of coal on unit investment cost

Table 12. LIMESTONE SLURRY PROCESS INVESTMENT WITH MODIFIED PROJECT SCOPE

	Investment, \$/kW
BASE INVESTMENT - LIMESTONE SLURRY PROCESS (Including Fly Ash Removal But Not Disposal)	
500-MW new coal-fired unit burning coal with 3.5% S, 12% ash, 90% SO ₂ removal, 30-year life 127,500 hours operation, onsite solids disposal, proven system, only pumps spared, no bypass ducts, experienced design and construction team, no overtime, 3-year program, 5% per year escalation, mid-1974 cost basis for scaling	50.30
A. Overtime to accelerate project or cover local demand requirements (50% of construction labor requirements)	3.20
B. Research and development costs for first of a kind process technology (as allowed by FPC accounting practice)	5.00
C. Power generation capital for lost capacity (normally covered by appropriate operating costs for power used in process)	4.50
D. Reliability provisions with added redundancy of scrubbers, other equipment, ducts and dampers, instrumentation for change over (assumes no permission to run power plant without meeting SO ₂ removal emission standards at all times)	7.00
E. Additional bypass ducts and dampers.	1.00
F. Retrofit difficulty--moderate, space available beyond stack, less than three shutdowns required for tie-ins, field fabrication feasible	10.00
G. Fly ash pond including closed-loop provisions	5.50
H. 500-ft stack added to project cost	6.00
I. Air quality monitoring system, 2-15 mile radius, 10 stations	0.70
J. Cost escalation of 10% per year instead of 5%	4.80
K. Possible delay of up to 2 years in equipment and material deliveries (1977 completion instead of 1975)	15.00
Total	113.00

Annual Operating Costs

For the annual operating costs covering 7000 hours per year operation, the following types of displays are used for each process:

Summarized effect of individual process areas on total costs-- presented for base case and existing 500-MW limestone case. See Tables 14 and 15.

Case variations--operating cost summaries presented for each of the 16 case variations. See Tables 16-21.

Table 22 presents the total average annual operating costs for the 16 case variations of all five processes excluding credit for any by-products produced. Corresponding to new 3.5% S coal-fired units, the relative ranking of average annual operating costs for the processes (base case) is as follows:

1. Limestone slurry
2. Lime slurry
3. Catalytic oxidation
4. Magnesia slurry - regeneration
5. Sodium solution - SO₂ reduction

Some important effects are not readily seen in comparing the ranking of processes for only one plant size and sulfur level. As determined in this study, the relative ranking of operating costs of the various processes is sensitive to changes in the assumed sulfur content of the fuel. As an illustration, a 500-MW, 1% S oil-fired power unit using the catalytic oxidation process has the highest operating cost of the five processes; but for fuel oils with sulfur contents greater than 4.0%, this process has the lowest cost. Although the sodium process is not the lowest cost process for any of the cases presented in this study, it becomes more competitive for low-sulfur fuel oil-fired units. Furthermore, at the expense of small additional investment, its operating cost could be lowered by designing the process with multiple-effect evaporators thereby reducing the overall energy requirements about 18%.

In all of the projected annual operating cost estimates, capital charges are the most significant components followed by maintenance, energy, and raw material costs in varying orders of importance. For the magnesia and sodium processes which require thermal energy for regeneration of absorbent, steam or fuel oil are the second most significant components. Lime cost (which requires thermal energy in preparation) is the second most predominant item for the lime slurry process, whereas maintenance is second in predominance for the limestone slurry and catalytic oxidation processes.

Table 14. LIMESTONE SLURRY PROCESS - TOTAL AVERAGE ANNUAL OPERATING COSTS
BASE CASE^a SUMMARY - AREA CONTRIBUTION ANALYSIS

	Total	Base materials handling	Feed preparation	Perkable scrubbing	SO ₂ scrubbing	Reheat	Feed drying	Utilities	Services	Construction	Total annual plant costs	Total annual operating costs	Percent of total annual costs
Direct capital investment, \$	10,049,000	419,000	899,000	3,201,000	4,741,000	554,000	854,000	19,131,000	17,000	419,000	100,000	100,000	0.29
Total capital investment, \$	24,183,000	419,000	899,000	3,201,000	4,741,000	554,000	854,000	19,131,000	17,000	419,000	100,000	100,000	0.29
Unit cost, \$/ton	171,000	171,000	171,000	171,000	171,000	171,000	171,000	171,000	171,000	171,000	171,000	171,000	0.29
Delivered raw material													
Limestone	4.60/ton												
Annual quantity, ton	171,000												
Annual cost, \$	171,000												
Subtotal raw material cost	171,000												
Construction costs													
Operating labor and expenses													
Annual quantity, man-hr													
Annual cost, \$													
Utilities													
Steam													
Annual quantity, M lb													
Annual cost, \$													
Process water													
Annual quantity, M gal													
Annual cost, \$													
Electricity													
Annual quantity, kWh													
Annual cost, \$													
Maintenance (labor and material)													
Percent of direct investment													
Annual cost, \$													
Air/fryer													
Annual quantity, lb													
Annual cost, \$													
Subtotal construction costs													
Subtotal direct costs													
Average capital charge at 14 % of capital investment													
Overhead													
Plant, 20% of construction costs													
Administrative, 10% of operating labor													
Subtotal indirect costs													
Total annual operating cost													
Percent of total annual operating cost													
Expenditure total and operating cost													
Expenditure total and operating cost													

^a100,000 tons coal-fired power plant, 2.5% S in fuel, 90% SO₂ removal, on-site solids disposal.
Expenditure total and operating cost, \$/ton.
Coal burned, 1,112,500 (tons/yr), 4,000 Btu/lb.
Steam generated, 1,112,500 (tons/yr), 4,000 Btu/lb.
Process gas output, 115 ft³/hr, 7,000 Btu/hr.
Maintenance (labor and material), 10% of operating labor.
Subtotal indirect costs, 10% of operating labor.
Subtotal indirect costs, 10% of operating labor.
Subtotal indirect costs, 10% of operating labor.

Table 15.

Sulfur removed: 13.4 M tons/yr, solids disposal: 210,000 tons/yr calcine solids including only hydrate water investment and operating cost for removal and disposal of fly ash excluded.

Table 16. LIMESTONE SLURRY PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	<u>Annual quantity</u>	<u>Unit cost, \$</u>	<u>Total annual cost, \$</u>	<u>Percent of total annual operating cost</u>
<u>Direct Costs</u>				
Delivered raw material				
Limestone	175.0 M tons	4.00/ton	700,000	9.09
Subtotal raw material			700,000	9.09
Conversion costs				
Operating labor and supervision	26,280 man-hr	8.00/man-hr	210,200	2.73
Utilities				
Steam	492,800 M lb	0.70/M lb	345,000	4.48
Process water	250,300 M gal	0.08/M gal	20,000	0.26
Electricity	78,740,000 kWh	0.010/kWh	787,400	10.22
Maintenance				
Labor and material, .08 x 16,069,000			1,285,500	16.69
Analyses			45,600	0.59
Subtotal conversion costs			2,693,700	34.97
Subtotal direct costs			3,393,700	44.06
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			3,749,300	48.68
Overhead				
Plant, 20% of conversion costs			538,700	6.99
Administrative, 10% of operating labor			21,000	0.27
Subtotal indirect costs			4,309,000	55.94
Total annual operating costs			7,702,700	100.00
	<u>Dollars/ton coal burned</u>	<u>Mills/kWh</u>	<u>Cents/million Btu heat input</u>	<u>Dollars/ton sulfur removed</u>
Equivalent unit operating cost	5.87	2.20	24.45	214.68

^a Basis:

Remaining life of power plant, 30 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$25,163,000; subtotal direct investment, \$16,069,000.
Working capital, \$572,600.
Investment and operating cost for disposal of fly ash excluded.

Table 17. LIME SLURRY PROCESS
VENTURI - VENTURI SCRUBBING SCHEME
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Lime	81.2 M tons	22.00/ton	1,786,400	22.05
Subtotal raw material			1,786,400	22.05
Conversion costs				
Operating labor and supervision	22,320 man-hr	8.00/man-hr	178,600	2.20
Utilities				
Steam	490,000 M lb	0.70/M lb	343,000	4.23
Process water	241,900 M gal	0.08/M gal	19,400	0.24
Electricity	74,100,000 kWh	0.010/kWh	741,000	9.15
Maintenance				
Labor and material, .08 x 14,318,000			1,145,400	14.14
Analyses			36,500	0.45
Subtotal conversion costs			2,463,900	30.41
Subtotal direct costs			4,250,300	52.46
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			3,340,900	41.24
Overhead				
Plant, 20% of conversion costs			492,800	6.08
Administrative, 10% of operating labor			17,900	0.22
Subtotal indirect costs			3,851,600	47.54
Total annual operating costs			8,101,900	100.00
	Dollars/ton coal burned	Mills/kWh	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	6.17	2.31	25.72	225.81

^a Basis:

Remaining life of power plant, 30 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$22,422,000; subtotal direct investment, \$14,318,000.
Working capital, \$744,000.
Investment and operating cost for disposal of fly ash excluded.

Table 18. LIME SLURRY PROCESS
VENTURI - VENTURI SCRUBBING SCHEME
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW existing coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Lime	83.0 M tons	22.00/ton	1,826,000	19.00
Subtotal raw material			1,826,000	19.00
Conversion costs				
Operating labor and supervision	22,320 man-hr	8.00/man-hr	178,600	1.86
Utilities				
Fuel oil (No. 6)	4,256,000 gal	0.23/gal	974,300	10.14
Process water	247,500 M gal	0.08/M gal	19,800	0.21
Electricity	67,310,000 kWh	0.010/kWh	673,100	7.00
Maintenance				
Labor and material, .08 x 15,913,000			1,273,000	13.23
Analyses			56,500	0.58
Subtotal conversion costs			3,155,300	32.62
Subtotal direct costs			4,981,300	51.62
<u>Indirect Costs</u>				
Average capital charges at 15.3% of total capital investment			3,982,100	41.42
Overhead				
Plant, 20% of conversion costs			631,100	6.57
Administrative, 10% of operating labor			17,900	0.19
Subtotal indirect costs			4,631,100	48.18
Total annual operating costs			9,612,400	100.00
	Dollars/ton coal burned	Cents/million Btu heat input	Dollars/ton sulfur removed	
Equivalent unit operating cost	7.16	2.75	29.85	262.06

^a

Notes:

- Remaining life of power plant, 25 yr.
- Coal burned, 1,341,700 tons/yr, 9,200 Btu/kWh.
- Stack gas reheat to 175°F.
- Power unit on-stream time, 7,000 hr/yr.
- Midwest plant location, 1976 operating costs.
- Total capital investment, \$26,027,000; subtotal direct investment, \$15,913,000.
- Working capital, \$865,500.
- Investment and operating cost for removal and disposal of fly ash excluded.

Table 19. MAGNESIA SLURRY - REGENERATION PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; 110,400 tons/yr 100% H₂SO₄)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Lime (1st stage neutralization)	134 tons	26.00/ton	3,500	0.04
Magnesium oxide (98%)	1,086 tons	155.00/ton	168,300	1.83
Coke	763 tons	15.00/ton	11,400	0.12
Catalyst	1,800 liters	1.65/liter	3,000	0.03
Subtotal raw material			186,200	2.02
Conversion costs				
Operating labor and supervision	39,200 man-hr	8.00/man-hr	313,600	3.40
Utilities				
Fuel oil (No. 6)	5,356,000 gal	0.23/gal	1,231,900	13.39
Steam	440,000 M lb	0.70/M lb	308,000	3.34
Heat credit	20,300 MM Btu	-0.60/MM Btu	(12,200)	(0.13)
Process water	2,207,500 M gal	0.04/M gal	88,300	0.96
Electricity	71,060,000 kWh	0.010/kWh	710,600	7.71
Maintenance				
Labor and material, .07 x 16,334,000			1,143,400	12.41
Analyses			102,000	1.11
Subtotal conversion costs			3,885,600	42.19
Subtotal direct costs			4,071,800	44.21
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			3,934,500	42.71
Overhead				
Plant, 20% of conversion costs			777,100	8.44
Administrative and marketing, 11% of conversion costs			427,400	4.64
Subtotal indirect costs			5,139,000	55.79
Total annual operating costs			9,210,800	100.00
	Dollars/ton 100% H ₂ SO ₄	Dollars/ton coal burned	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	83.43	7.02	29.24	255.43

^a Basis:

Remaining life of power plant, 30 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$26,406,000; subtotal direct investment, \$16,334,000.
Working capital, \$721,000.
Investment and operating cost for disposal of fly ash excluded.

Table 20. SODIUM SOLUTION - SO₂ REDUCTION PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; 32,700 tons/yr sulfur)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
<u>Delivered raw materials</u>				
Lime (1st stage neutralization)	154.1 tons	26.00/ton	3,500	0.03
Soda ash	9,300 tons	52.00/ton	483,600	4.17
Antioxidant	317,100 lb	2.00/lb	634,200	5.47
Catalyst			12,000	0.10
Subtotal raw materials			1,133,300	9.77
<u>Conversion costs</u>				
Operating labor and supervision	46,500 man-hr	8.00/man-hr	372,000	3.21
<u>Utilities</u>				
Natural gas	509,500 mcf	1.00/mcf	509,500	4.39
Steam	2,137,800 M lb	0.70/M lb	1,496,500	12.90
Heat credit	62,900 MM Btu	-0.60/MM Btu	(37,700)	(0.33)
Process water	9,933,400 M gal	0.00/M gal	199,100	1.72
Electricity	74,190,000 kWh	0.010/kWh	741,900	6.39
Maintenance				
Labor and material, .06 x 18,861,000			1,131,700	9.75
Analyses			102,900	0.90
Subtotal conversion costs			4,522,900	38.98
Subtotal direct costs			5,656,200	48.75
<u>Indirect Costs</u>				
<u>Average capital charges at 14.0% of total capital investment</u>				
Overhead			4,543,200	39.16
<u>Plant, 20% of conversion costs</u>				
Administrative and marketing			904,600	7.80
Subtotal indirect costs			477,500	4.29
			5,945,300	51.25
Total annual operating costs			11,601,500	100.00
	Dollars/ton Product sulfur	Dollars/ton coal burned	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	354.79	8.84	3.31	323.34

^a Basis:

- Remaining life of power plant, 30 yr.
- Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
- Stack gas reheat to 175°F.
- Power unit on-stream time, 7,000 hr/yr.
- Midwest plant location, 1975 operating costs.
- Total capital investment, \$30,491,000; subtotal direct investment, \$18,861,000.
- Working capital, \$1,015,000.
- Investment and operating cost for disposal of fly ash excluded.

Table 21. CATALYTIC OXIDATION PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^B

(400-MW new, coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; 104,400 tons/yr 100% H₂SO₄)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Catalyst	104,700 liters	1.65/liter	172,800	1.95
Subtotal raw material			172,800	1.95
Conversion costs				
Operating labor and supervision	7,840 man-hr	8.00/man-hr	63,100	0.71
Utilities				
Steam	179,000 M lb	0.70/M lb	125,300	1.41
Heat credit	987,000 MM Btu	-0.60/MM Btu	(592,200)	(6.67)
Process water	512,000 M gal	0.06/M gal	25,000	0.28
Electricity	90,440,000 kWh	0.010/kWh	904,400	10.19
Maintenance				
Labor and material, 1.04 x 25,468,000			1,014,700	11.45
Analyses			43,000	0.54
Subtotal conversion costs			1,533,300	17.39
Subtotal direct costs			1,761,100	19.84
<u>Indirect Costs</u>				
Average capital charges at 14.04 of total capital investment			6,367,700	71.76
Overhead				
Plant, 20% of conversion costs			317,700	3.59
Administrative and marketing			427,400	4.82
Subtotal indirect costs			7,112,800	80.16
Total annual operating costs			8,873,900	100.00
	Dollars/ton 100% H ₂ SO ₄	Dollars/ton coal burned	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	80.75	6.76	25.17	247.52

^B Basis:

Remaining life of power plant, 50 yr.
Coal burned, 1,512,000 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 170°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 100% operating costs.
Total capital investment, \$62,756,000; subtotal direct investment, \$4,568,000.
Working capital, \$505,700.
Investment and operating cost for disposal of fly ash excluded.

Table 22. SUMMARY - TOTAL AVERAGE ANNUAL OPERATING COSTS^a

Excluding Credit for Byproducts

Case	Years life	Limestone process			Gypsum process			Sodium process			Cat-Ox process		
		Total annual operating cost, \$	Mille/ kwh	Total annual operating cost, \$	Mille/ kwh	Total annual operating cost, \$	Mille/ kwh	Total annual operating cost, \$	Mille/ kwh	Total annual operating cost, \$	Mille/ kwh		
Coal-fired power unit													
2x4 3/2 removal; onsite solids disposal	30	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
200 Mw new, 3.5% sulfur	30	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
200 Mw existing, 3.5% sulfur	20	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
500 Mw new, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
500 Mw existing, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
1000 Mw new, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
1000 Mw existing, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
2x4 3/2 removal; offsite solids disposal	30	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
200 Mw new, 3.5% sulfur	30	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
200 Mw existing, 3.5% sulfur	20	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
500 Mw new, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
500 Mw existing, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
1000 Mw new, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
1000 Mw existing, 3.5% sulfur	30	7,378,500	2.26	9,612,500	2.75	9,607,300	2.75	14,663,000	4.19	12,599,600			3.54
2x4 3/2 removal (existing unit without existing particulate collection facilities)	25	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
500 Mw existing, 3.5% sulfur	25	3,301,500	2.30	4,161,300	2.27	4,776,300	3.41	5,071,700	4.27	4,258,700			3.02
Oil-fired power unit													
2x4 3/2 removal; onsite solids disposal	30	2,342,000	2.05	3,411,500	2.44	3,204,400	2.29	4,264,200	3.05	2,750,100			1.96
200 Mw new, 2.5% sulfur	30	2,342,000	2.05	3,411,500	2.44	3,204,400	2.29	4,264,200	3.05	2,750,100			1.96
200 Mw existing, 2.5% sulfur	20	2,342,000	2.05	3,411,500	2.44	3,204,400	2.29	4,264,200	3.05	2,750,100			1.96
500 Mw new, 2.5% sulfur	30	5,564,400	1.59	6,352,300	1.96	6,092,700	1.74	9,505,100	2.37	5,677,500			1.62
500 Mw existing, 2.5% sulfur	30	5,564,400	1.59	6,352,300	1.96	6,092,700	1.74	9,505,100	2.37	5,677,500			1.62
1000 Mw new, 2.5% sulfur	30	5,564,400	1.59	6,352,300	1.96	6,092,700	1.74	9,505,100	2.37	5,677,500			1.62
1000 Mw existing, 2.5% sulfur	30	5,564,400	1.59	6,352,300	1.96	6,092,700	1.74	9,505,100	2.37	5,677,500			1.62
2x4 3/2 removal (existing unit without existing particulate collection facilities)	25	2,342,000	2.05	3,411,500	2.44	3,204,400	2.29	4,264,200	3.05	2,750,100			1.96
500 Mw existing, 2.5% sulfur	25	2,342,000	2.05	3,411,500	2.44	3,204,400	2.29	4,264,200	3.05	2,750,100			1.96
1000 Mw new, 2.5% sulfur	30	5,564,400	1.59	6,352,300	1.96	6,092,700	1.74	9,505,100	2.37	5,677,500			1.62
1000 Mw existing, 2.5% sulfur	30	5,564,400	1.59	6,352,300	1.96	6,092,700	1.74	9,505,100	2.37	5,677,500			1.62

^a Power unit on-stream time, 7000 hr/yr. Midwest plant location, 17% operating costs. Investment and operating cost for disposal of fly ash excluded.

The effect on annual operating costs of designing for 80% SO₂ removal as compared to 90% is relatively small. Operating cost savings range from 3.6 to 4.2% of the projected operating costs at 90% removal for the limestone and lime slurry processes.

The effect of excluding product credits from the annual operating cost estimates is illustrated in Table 23 which compares the projected annual base case operating costs for the five processes with the revenue from sale of byproducts.

Table 23. COMPARISON OF PROJECTED ANNUAL OPERATING COSTS AND PRODUCT CREDITS FOR BASE CASE ESTIMATES AT 7000 HOURS ANNUAL OPERATION

Process	Total average annual operating cost, \$	Annual credit for byproducts, \$ ^a	Net annual operating cost, \$
Limestone	7,702,700	-	7,702,700
Lime	8,101,900	-	8,101,900
Magnesia	9,210,800	883,200	8,327,600
Sodium	11,601,500	1,077,500	10,524,000
Cat-Ox	8,873,900	659,400	8,214,500

^a Corresponds to credit of \$8/ton 100% H₂SO₄ as 98% H₂SO₄ for magnesia slurry process; \$25/short ton sulfur, \$20/ton Na₂SO₄ for the sodium solution process; and \$6/ton 100% H₂SO₄ as 80% H₂SO₄ for the catalytic oxidation process.

Lifetime Operating Cost

Assumed credits for salable byproducts are reflected in the lifetime operating costs. The lifetime operating costs are based on a prescribed 30-year operating profile--7000 hours the first 10 years; 5000 hours the next 5 years; 3500 hours the next 5 years; 1500 hours the last 10 years (for examples see Tables 24-25). Table 26 presents a summary of the cumulative discounted process costs and equivalent levelized unit increase (decrease) in the cost of power over the life of the power unit for the five processes. Table 27 shows the revenue assumed for each product and the actual and discounted cumulative credit for products which are included in the cumulative discounted process costs for the base case. Figures 14 and 15 show the effect of power unit size and sulfur content of coal on the levelized unit operating cost. With some exceptions, the relative ranking and trends in levelized unit operating costs are somewhat similar to those projected and discussed for annual operating costs. Lifetime levelized unit operating costs are slightly higher than corresponding average annual unit operating costs because of the declining operating profile of the power unit.

Table 24. LIMESTONE SLURRY PROCESS, 500-MW NEW COAL-FIRED POWER UNIT,
3.5% SULFUR IN FUEL, 90% SO₂ REMOVAL, REGULATED COMPANY ECONOMICS

YEARS ANNUAL POWER OPERA- TION UNIT START	YEARS ANNUAL POWER UNIT REQUIREMENT, CONSUMPTION, MILLION BTU /YEAR	FUEL POWER UNIT TONS COAL /YEAR	SULFUR REMOVED BY FUELS CONTROL PROCESS, TMS/YEAR	BY-PRODUCT EQUIVALENT TMS/YEAR	NET REVENUE, \$/TOM	TOTAL INVESTMENT, \$	TOTAL INVESTMENT REGULATED FOR POWER COMPANY, \$/YEAR	TOTAL SALES \$/YEAR	NET ANNUAL INCREASE (DECREASE) IN COST OF POWER, \$/TOM	CUMULATIVE NET INCREASE (DECREASE) IN COST OF POWER, \$/TOM
1	7000	3150000	131250	20000	0.0	10320500	0	10320500	0	10320500
2	7000	3150000	131250	20000	0.0	10146000	0	10146000	0	20466500
3	7000	3150000	131250	20000	0.0	9971000	0	9971000	0	30437500
4	7000	3150000	131250	20000	0.0	9796000	0	9796000	0	40398500
5	7000	3150000	131250	20000	0.0	9621000	0	9621000	0	50359500
6	7000	3150000	131250	20000	0.0	9446000	0	9446000	0	60320500
7	7000	3150000	131250	20000	0.0	9271000	0	9271000	0	70281500
8	7000	3150000	131250	20000	0.0	9096000	0	9096000	0	80242500
9	7000	3150000	131250	20000	0.0	8921000	0	8921000	0	90203500
10	7000	3150000	131250	20000	0.0	8746000	0	8746000	0	100164500
11	5000	2250000	93750	14700	0.0	7640000	0	7640000	0	107804500
12	5000	2250000	93750	14700	0.0	7465000	0	7465000	0	115444500
13	5000	2250000	93750	14700	0.0	7290000	0	7290000	0	123084500
14	5000	2250000	93750	14700	0.0	7115000	0	7115000	0	130724500
15	5000	2250000	93750	14700	0.0	6940000	0	6940000	0	138364500
16	3500	1575000	65625	10300	0.0	6229000	0	6229000	0	144593500
17	3500	1575000	65625	10300	0.0	6054000	0	6054000	0	150822500
18	3500	1575000	65625	10300	0.0	5879000	0	5879000	0	157051500
19	3500	1575000	65625	10300	0.0	5704000	0	5704000	0	163280500
20	3500	1575000	65625	10300	0.0	5529000	0	5529000	0	169509500
21	3500	1575000	65625	10300	0.0	5354000	0	5354000	0	175738500
22	3500	1575000	65625	10300	0.0	5179000	0	5179000	0	181967500
23	3500	1575000	65625	10300	0.0	5004000	0	5004000	0	188196500
24	3500	1575000	65625	10300	0.0	4829000	0	4829000	0	194425500
25	3500	1575000	65625	10300	0.0	4654000	0	4654000	0	200654500
26	3500	1575000	65625	10300	0.0	4479000	0	4479000	0	206883500
27	3500	1575000	65625	10300	0.0	4304000	0	4304000	0	213112500
28	3500	1575000	65625	10300	0.0	4129000	0	4129000	0	219341500
29	3500	1575000	65625	10300	0.0	3954000	0	3954000	0	225570500
30	3500	1575000	65625	10300	0.0	3779000	0	3779000	0	231799500
TOTAL 127500 57375000 2304000 65300 3751500 19310500 0 19310500										
LIFETIME AVERAGE INCREASE IN UNIT OPERATING COST										
DOLLARS PER TON OF COAL BURNED										
MILLS PER MILLION BTU HEAT INPUT										
CENTS PER MILLION BTU HEAT INPUT										
DOLLARS PER TON OF SULFUR REMOVED										
CENTS PER TON OF SULFUR REMOVED										
DOLLARS PER TON OF COAL BURNED										
CENTS PER TON OF COAL BURNED										
DOLLARS PER TON OF SULFUR REMOVED										
CENTS PER TON OF SULFUR REMOVED										

Table 25. LIMESTONE SLURRY PROCESS, 500-MW EXISTING COAL-FIRED POWER UNIT,
3.5% SULFUR IN FUEL, 90% SO₂ REMOVAL, REGULATED COMPANY ECONOMICS

YEARS AFTER POWER UNIT START	ANNUAL HEAT REQUIREMENT, MMBTU/HR	FUEL CONSUMPTION, TONS COAL /YEAR	SULFUR REMOVED BY POLLUTION CONTROL PROCESS, TONS/YEAR	PI-PRODUCT RATE, TONS/YEAR	NET REVENUE, \$/TON	WASTE SOLIDS \$/YEAR	TOTAL INVESTMENT \$	TOTAL COST INCLUDING REGULATED EMISSIONS \$/YEAR	NET ANNUAL INCREASE (DECREASE) IN COST OF POWER, \$/MWH	CUMULATIVE NET INCREASE (DECREASE) IN COST OF POWER, \$/MWH
1	7000	3220000	1341700	212600	0.0	0.0	10291700	0	10291700	10291700
2	7000	3220000	1341700	212600	0.0	0.0	10101600	0	10101600	20393200
3	7000	3220000	1341700	212600	0.0	0.0	9911500	0	9911500	30494700
4	7000	3220000	1341700	212600	0.0	0.0	9721400	0	9721400	40596200
5	7000	3220000	1341700	212600	0.0	0.0	9531300	0	9531300	50697700
6	7000	3220000	1341700	212600	0.0	0.0	9341200	0	9341200	60799200
7	7000	3220000	1341700	212600	0.0	0.0	9151100	0	9151100	70900700
8	7000	3220000	1341700	212600	0.0	0.0	8961000	0	8961000	81002200
9	7000	3220000	1341700	212600	0.0	0.0	8770900	0	8770900	91103700
10	7000	3220000	1341700	212600	0.0	0.0	8580800	0	8580800	101205200
11	7000	3220000	1341700	212600	0.0	0.0	8390700	0	8390700	111306700
12	7000	3220000	1341700	212600	0.0	0.0	8200600	0	8200600	121408200
13	7000	3220000	1341700	212600	0.0	0.0	8010500	0	8010500	131509700
14	7000	3220000	1341700	212600	0.0	0.0	7820400	0	7820400	141611200
15	7000	3220000	1341700	212600	0.0	0.0	7630300	0	7630300	151712700
16	7000	3220000	1341700	212600	0.0	0.0	7440200	0	7440200	161814200
17	7000	3220000	1341700	212600	0.0	0.0	7250100	0	7250100	171915700
18	7000	3220000	1341700	212600	0.0	0.0	7060000	0	7060000	182017200
19	7000	3220000	1341700	212600	0.0	0.0	6869900	0	6869900	192118700
20	7000	3220000	1341700	212600	0.0	0.0	6679800	0	6679800	202220200
21	7000	3220000	1341700	212600	0.0	0.0	6489700	0	6489700	212321700
22	7000	3220000	1341700	212600	0.0	0.0	6299600	0	6299600	222423200
23	7000	3220000	1341700	212600	0.0	0.0	6109500	0	6109500	232524700
24	7000	3220000	1341700	212600	0.0	0.0	5919400	0	5919400	242626200
25	7000	3220000	1341700	212600	0.0	0.0	5729300	0	5729300	252727700
26	7000	3220000	1341700	212600	0.0	0.0	5539200	0	5539200	262829200
27	7000	3220000	1341700	212600	0.0	0.0	5349100	0	5349100	272930700
28	7000	3220000	1341700	212600	0.0	0.0	5159000	0	5159000	283032200
29	7000	3220000	1341700	212600	0.0	0.0	4968900	0	4968900	293133700
30	7000	3220000	1341700	212600	0.0	0.0	4778800	0	4778800	303235200
31	7000	3220000	1341700	212600	0.0	0.0	4588700	0	4588700	313336700
32	7000	3220000	1341700	212600	0.0	0.0	4398600	0	4398600	323438200
33	7000	3220000	1341700	212600	0.0	0.0	4208500	0	4208500	333539700
34	7000	3220000	1341700	212600	0.0	0.0	4018400	0	4018400	343641200
35	7000	3220000	1341700	212600	0.0	0.0	3828300	0	3828300	353742700
36	7000	3220000	1341700	212600	0.0	0.0	3638200	0	3638200	363844200
37	7000	3220000	1341700	212600	0.0	0.0	3448100	0	3448100	373945700
38	7000	3220000	1341700	212600	0.0	0.0	3258000	0	3258000	384047200
39	7000	3220000	1341700	212600	0.0	0.0	3067900	0	3067900	394148700
40	7000	3220000	1341700	212600	0.0	0.0	2877800	0	2877800	404250200
41	7000	3220000	1341700	212600	0.0	0.0	2687700	0	2687700	414351700
42	7000	3220000	1341700	212600	0.0	0.0	2497600	0	2497600	424453200
43	7000	3220000	1341700	212600	0.0	0.0	2307500	0	2307500	434554700
44	7000	3220000	1341700	212600	0.0	0.0	2117400	0	2117400	444656200
45	7000	3220000	1341700	212600	0.0	0.0	1927300	0	1927300	454757700
46	7000	3220000	1341700	212600	0.0	0.0	1737200	0	1737200	464859200
47	7000	3220000	1341700	212600	0.0	0.0	1547100	0	1547100	474960700
48	7000	3220000	1341700	212600	0.0	0.0	1357000	0	1357000	485062200
49	7000	3220000	1341700	212600	0.0	0.0	1166900	0	1166900	495163700
50	7000	3220000	1341700	212600	0.0	0.0	976800	0	976800	505265200
51	7000	3220000	1341700	212600	0.0	0.0	786700	0	786700	515366700
52	7000	3220000	1341700	212600	0.0	0.0	596600	0	596600	525468200
53	7000	3220000	1341700	212600	0.0	0.0	406500	0	406500	535569700
54	7000	3220000	1341700	212600	0.0	0.0	216400	0	216400	545671200
55	7000	3220000	1341700	212600	0.0	0.0	26300	0	26300	555772700
56	7000	3220000	1341700	212600	0.0	0.0	0	0	0	565874200
57	7000	3220000	1341700	212600	0.0	0.0	0	0	0	575975700
58	7000	3220000	1341700	212600	0.0	0.0	0	0	0	586077200
59	7000	3220000	1341700	212600	0.0	0.0	0	0	0	596178700
60	7000	3220000	1341700	212600	0.0	0.0	0	0	0	606280200
61	7000	3220000	1341700	212600	0.0	0.0	0	0	0	616381700
62	7000	3220000	1341700	212600	0.0	0.0	0	0	0	626483200
63	7000	3220000	1341700	212600	0.0	0.0	0	0	0	636584700
64	7000	3220000	1341700	212600	0.0	0.0	0	0	0	646686200
65	7000	3220000	1341700	212600	0.0	0.0	0	0	0	656787700
66	7000	3220000	1341700	212600	0.0	0.0	0	0	0	666889200
67	7000	3220000	1341700	212600	0.0	0.0	0	0	0	676990700
68	7000	3220000	1341700	212600	0.0	0.0	0	0	0	687092200
69	7000	3220000	1341700	212600	0.0	0.0	0	0	0	697193700
70	7000	3220000	1341700	212600	0.0	0.0	0	0	0	707295200
71	7000	3220000	1341700	212600	0.0	0.0	0	0	0	717396700
72	7000	3220000	1341700	212600	0.0	0.0	0	0	0	727498200
73	7000	3220000	1341700	212600	0.0	0.0	0	0	0	737599700
74	7000	3220000	1341700	212600	0.0	0.0	0	0	0	747701200
75	7000	3220000	1341700	212600	0.0	0.0	0	0	0	757802700
76	7000	3220000	1341700	212600	0.0	0.0	0	0	0	767904200
77	7000	3220000	1341700	212600	0.0	0.0	0	0	0	778005700
78	7000	3220000	1341700	212600	0.0	0.0	0	0	0	788107200
79	7000	3220000	1341700	212600	0.0	0.0	0	0	0	798208700
80	7000	3220000	1341700	212600	0.0	0.0	0	0	0	808310200
81	7000	3220000	1341700	212600	0.0	0.0	0	0	0	818411700
82	7000	3220000	1341700	212600	0.0	0.0	0	0	0	828513200
83	7000	3220000	1341700	212600	0.0	0.0	0	0	0	838614700
84	7000	3220000	1341700	212600	0.0	0.0	0	0	0	848716200
85	7000	3220000	1341700	212600	0.0	0.0	0	0	0	858817700
86	7000	3220000	1341700	212600	0.0	0.0	0	0	0	868919200
87	7000	3220000	1341700	212600	0.0	0.0	0	0	0	879020700
88	7000	3220000	1341700	212600	0.0	0.0	0	0	0	889122200
89	7000	3220000	1341700	212600	0.0	0.0	0	0	0	899223700
90	7000	3220000	1341700	212600	0.0	0.0	0	0	0	909325200
91	7000	3220000	1341700	212600	0.0	0.0	0	0	0	919426700
92	7000	3220000	1341700	212600	0.0	0.0	0	0	0	929528200
93	7000	3220000	1341700	212600	0.0	0.0	0	0	0	939629700
94	7000	3220000	1341700	212600	0.0	0.0	0	0	0	949731200
95	7000	3220000	1341700	212600	0.0	0.0	0	0	0	959832700
96	7000	3220000	1341700	212600	0.0	0.0	0	0	0	969934200
97	7000	3220000	1341700	212600	0.0	0.0	0	0	0	980035700
98	7000	3220000	1341700	212600	0.0	0.0	0	0	0	990137200
99	7000	3220000	1341700	212600	0.0	0.0	0	0	0	1000238700
100	7000	3220000	1341700	212600	0.0	0.0	0	0	0	1010140200

Table 26. SUMMARY - CUMULATIVE DISCOUNTED PROCESS COSTS AND EQUIVALENT LEVELIZED UNIT INCREASE (DECREASE) IN COST OF POWER OVER THE LIFE OF THE POWER UNIT^a

Including Credit for Byproducts

Case	Limestone process				Lime process				Magnesia process				Sodium process				Cat-Ox process			
	Cumulative present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Cumulative present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Cumulative present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Cumulative present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Cumulative present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c	Present worth net increase in unit operating cost, mills/kwh ^c	Levelized increase (decrease) in unit operating cost, mills/kwh ^c
Coal-fired power unit																				
90% SO ₂ removal; onsite solids disposal																				
200 Mw new, 3.5% sulfur	40,142,800	3.66	41,112,500	3.75	44,860,100	4.09	55,045,000	5.02	64,823,500	4.08	64,823,500	4.08	55,045,000	5.02	64,823,500	4.08	64,823,500	4.08	55,045,000	5.02
200 Mw existing, 3.5% sulfur	29,067,800	4.75	34,379,000	5.69	36,106,200	5.87	48,168,300	7.90	42,423,000	6.90	42,423,000	6.90	48,168,300	7.90	42,423,000	6.90	42,423,000	6.90	48,168,300	7.90
500 Mw new, 3.5% sulfur	76,550,000	3.09	84,117,600	3.69	75,892,200	3.43	113,285,500	5.90	106,607,900	4.67	106,607,900	4.67	113,285,500	5.90	106,607,900	4.67	106,607,900	4.67	113,285,500	5.90
500 Mw existing, 3.5% sulfur	69,114,200	2.53	66,709,000	2.60	71,503,600	2.61	95,604,500	3.42	99,150,500	3.44	99,150,500	3.44	95,604,500	3.42	99,150,500	3.44	99,150,500	3.44	95,604,500	3.42
500 Mw new, 5.0% sulfur	76,439,900	2.96	79,593,300	2.90	84,249,500	3.07	104,236,300	3.90	94,350,500	3.44	94,350,500	3.44	104,236,300	3.90	94,350,500	3.44	94,350,500	3.44	104,236,300	3.90
500 Mw existing, 5.0% sulfur	56,426,800	3.15	89,293,900	3.25	96,621,900	3.49	121,660,300	4.43	92,350,500	3.43	92,350,500	3.43	121,660,300	4.43	92,350,500	3.43	92,350,500	3.43	121,660,300	4.43
1000 Mw new, 3.5% sulfur	111,285,400	2.45	130,977,300	2.87	121,156,200	2.66	199,466,400	4.13	181,012,700	3.97	181,012,700	3.97	199,466,400	4.13	181,012,700	3.97	181,012,700	3.97	199,466,400	4.13
1000 Mw existing, 3.5% sulfur	120,015,500	2.19	121,799,900	2.22	126,808,000	2.31	160,375,200	2.92	149,117,600	2.70	149,117,600	2.70	160,375,200	2.92	149,117,600	2.70	149,117,600	2.70	160,375,200	2.92
80% SO ₂ removal; onsite solids disposal																				
500 Mw new, 3.5% sulfur	75,259,300	2.74	76,687,900	2.80	81,119,900	2.96	98,245,000	3.59	88,245,000	3.59	88,245,000	3.59	98,245,000	3.59	88,245,000	3.59	88,245,000	3.59	98,245,000	3.59
90% SO ₂ removal; offsite solids disposal																				
500 Mw new, 3.5% sulfur	80,426,200	2.93	80,903,300	2.95	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
90% SO ₂ removal (existing unit without existing particulate collection facilities)																				
500 Mw existing, 3.5% sulfur	57,145,300	3.82	84,984,200	3.72	93,875,800	4.12	130,713,900	5.73	119,124,800	5.22	119,124,800	5.22	130,713,900	5.73	119,124,800	5.22	119,124,800	5.22	130,713,900	5.73
Oil-fired power unit																				
90% SO ₂ removal; onsite solids disposal																				
200 Mw new, 2.5% sulfur	28,291,000	2.58	33,612,500	3.06	30,089,900	2.74	39,147,900	3.57	29,653,900	2.70	29,653,900	2.70	39,147,900	3.57	29,653,900	2.70	29,653,900	2.70	39,147,900	3.57
500 Mw new, 1.0% sulfur	46,404,800	1.69	56,505,800	2.06	44,030,300	1.61	55,025,400	2.01	43,574,000	2.32	43,574,000	2.32	55,025,400	2.01	43,574,000	2.32	43,574,000	2.32	55,025,400	2.01
500 Mw new, 2.5% sulfur	54,743,300	2.00	66,727,200	2.43	55,673,400	2.03	74,204,600	2.70	61,591,500	2.45	61,591,500	2.45	74,204,600	2.70	61,591,500	2.45	61,591,500	2.45	74,204,600	2.70
500 Mw new, 4.0% sulfur	61,808,400	2.25	74,963,500	2.73	65,572,800	2.39	91,687,900	3.35	59,249,500	2.16	59,249,500	2.16	91,687,900	3.35	59,249,500	2.16	59,249,500	2.16	91,687,900	3.35
500 Mw existing, 2.5% sulfur	58,559,500	2.56	70,126,200	3.07	61,335,500	3.07	82,592,100	3.63	58,505,200	2.42	58,505,200	2.42	82,592,100	3.63	58,505,200	2.42	58,505,200	2.42	82,592,100	3.63
1000 Mw new, 2.5% sulfur	97,271,700	1.59	105,411,900	1.88	85,962,400	1.57	115,705,700	2.16	85,955,500	1.77	85,955,500	1.77	115,705,700	2.16	85,955,500	1.77	85,955,500	1.77	115,705,700	2.16

^a Basis:

Over previously defined power unit operating profile. 30 yr life; 7000 hr - 10 yr, 5000 hr - 5 yr, 3500 hr - 10 yr.

Midwest plant location, 1975 operating costs.

Investment and operating cost for disposal of fly ash excluded.

Constant labor cost assumed over life of project.

^b Discounted at 10% to initial year.

^c Equivalent to discounted process cost over life of power unit.

Table 27. LIFETIME BYPRODUCT PRODUCTION AND CREDIT

Base Case, 500-MW New, Coal-Fired, 3.5% S, 30 Yrs Remaining Life

Process	Equivalent lifetime production		Net revenue, \$/short ton	Cumulative revenue	
	Product	Short tons		Actual \$	Dis- counted \$
Magnesia slurry - regeneration	100% H ₂ SO ₄ ^a	2,011,500	8.00	16,092,000	6,923,300
Sodium solution - SO ₂ reduction	Sulfur	595,000	25.00		
	Na ₂ SO ₄	237,000	20.00	19,627,500	8,446,300
Catalytic oxidation	100% H ₂ SO ₄ ^b	2,002,500	6.00	12,015,000	5,168,900

^a As 93% H₂SO₄.^b As 80% H₂SO₄.

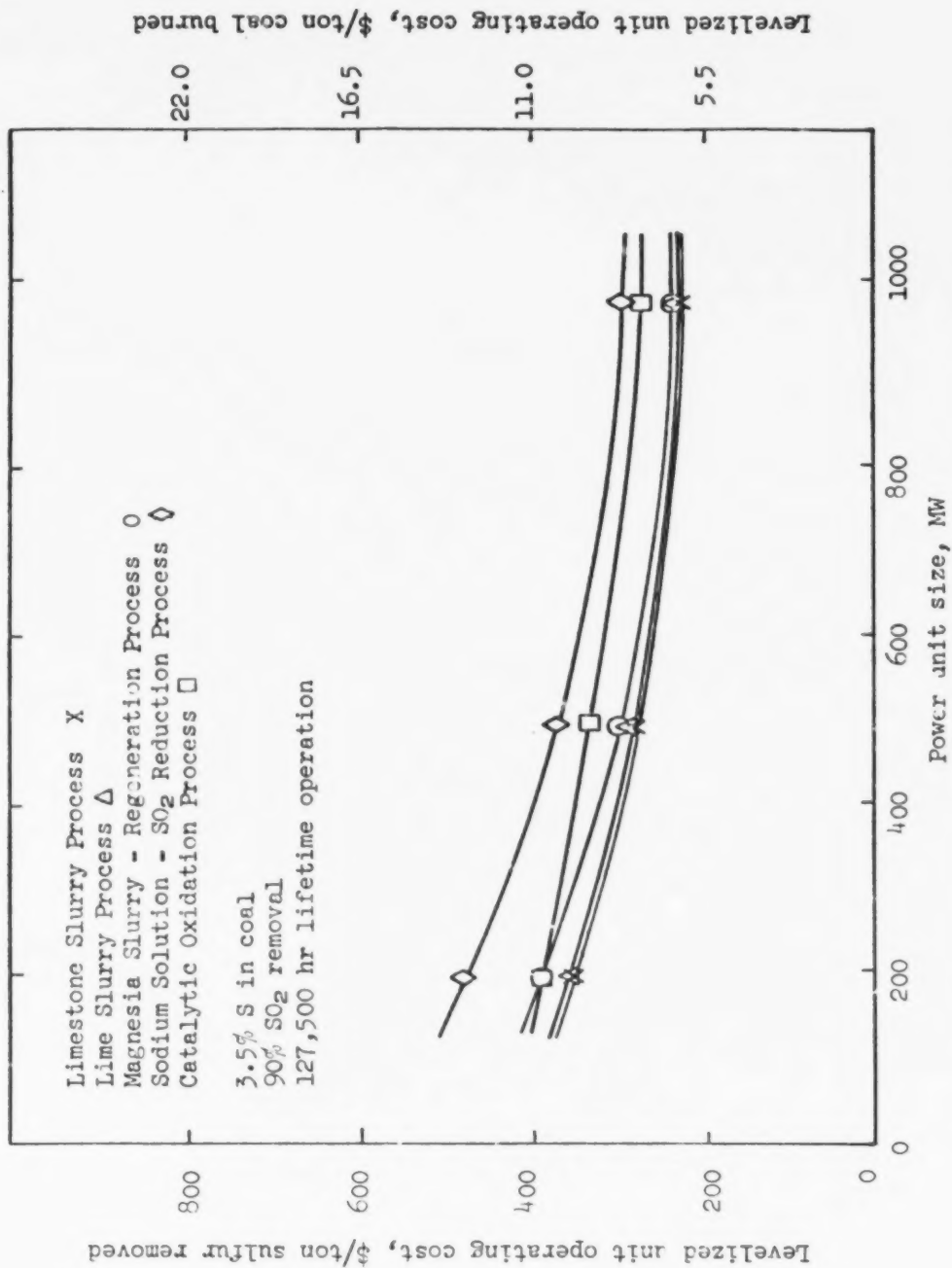


Figure 14. New coal-fired units - the effect of power unit size on leveled unit operating cost - regulated economics

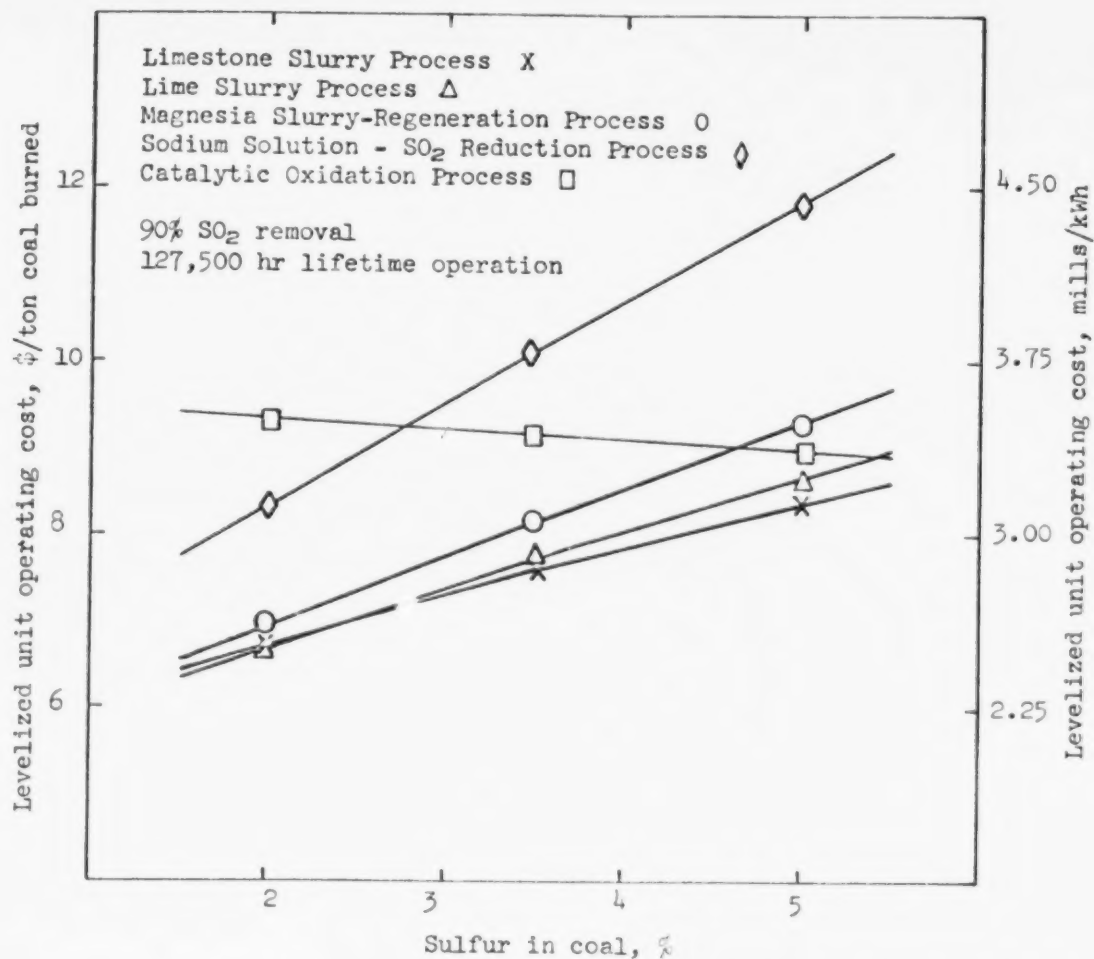


Figure 15. 500-MW new coal-fired units - effect of sulfur content of coal on levelized unit operating cost - regulated economics

TVA IN-HOUSE ESTIMATES

In addition to the more detailed estimates prepared for EPA, a series of in-house studies have also been undertaken by TVA to determine the attractiveness of several process options. These include limestone scrubbing using benzoic acid as a slurry dissolution promoter; lime scrubbing using a venturi - mobile-bed contactor rather than two venturi stages; lime scrubbing including a limestone calcination unit; sodium scrubbing with sulfuric acid production rather than sulfur; and two double-alkali systems--one using ammonia scrubbing, the other sodium scrubbing. Because the data available for the double-alkali process evaluations were limited, the accuracy of these estimates is less than the others.

Seven TVA case estimates were prepared. When evaluating lime scrubbing with a mobile-bed contactor, both a new and a retrofit case (500 MW, 3.5% S in coal) were estimated to determine the merit of using such a single-stage device for SO₂ removal. For the other five process options, only new systems were evaluated. Except for specific process requirements, the same criteria used to prepare the EPA estimates were used for these evaluations. The summarized investment and operating cost results are given in Table 28. Additional comparative details are given in Tables 29-41.

CONCLUSIONS

Conclusions derived from the EPA-TVA detailed study results are as follows:

Investment

1. For coal-fired systems, the lime scrubbing process has the lowest investment and catalytic oxidation has the highest. Because lime process definition specifies venturi scrubbers and two stages are required for SO₂ removal, the limestone and magnesia processes become comparatively less expensive on oil-fired units; limestone investment is the lowest and magnesia investment is very close. All four wet scrubbing processes are within 21% of each other.
2. As sulfur content of fuel increases, the relative investment ranking of the wet scrubbing processes changes; lime is still lowest for coal-fired units but for oil-fired units, limestone or magnesia investments are the lowest. Although these investments are very close throughout the range, magnesia investment is slightly lower for the low-sulfur oil, and limestone investment is slightly lower for the medium- and high-sulfur oils.

Table 28. INVESTMENT AND OPERATING COST RESULTS FROM CVA IN-HOUSE STUDIES^a

Process	Years life	Capital investment \$	\$/kW	Average annual operating costs \$	mills/kWh
Limestone-benzoic acid slurry process ^b 500 MW new 3.5% sulfur	30	23,473,000	46.9	7,439,000	2.13
Lime slurry process; venturi-mobile bed scrubbing scheme ^b 500 MW new 3.5% sulfur	30	23,363,000	46.7	8,146,000	2.33
Lime slurry process; mobile bed scrubbing scheme ^b 500 MW existing, ^c 3.5% sulfur	25	20,327,000	40.7	8,270,600	2.36
Lime slurry process; venturi - venturi scrubbing scheme with onsite calcination ^b 500 MW new, 3.5% sulfur	30	27,452,000	54.9	8,101,900 ^d	2.31 ^d
Sodium solution - H ₂ SO ₄ production process 500 MW new, 3.5% sulfur	30	30,138,000	60.3	11,054,800	3.16
Ammonia-limestone/lime double-alkali process ^b 500 MW new 3.5% sulfur	30	31,087,000	62.2	9,669,300	2.76
Sodium-limestone double-alkali process ^b 500 MW new 3.5% sulfur	30	30,759,000	61.5	9,545,300	2.73

^a Stack gas reheat to 175°F. Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974. Operating costs based on power unit on-stream time of 7000 hours; costs correspond to 1975. Minimum in process storage; only pumps are spared. Investment and operating cost for disposal of fly ash excluded.

^b Construction labor shortages with accompanying overtime pay incentive not considered.

^c Limestone raw material cost, \$4/ton; lime raw material cost, \$22.00/ton. Onsite disposal pond located one mile from power plant.

^d Assumes 98.7% effective electrostatic precipitator already installed.

^e Same as operating cost for lime slurry process in EPA study, since EPA study utilized estimated production cost to establish purchased cost of lime.

Table 29. LIMESTONE - BENZOIC ACID SLURRY PROCESS
^a
SUMMARY OF ESTIMATED FIXED INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Investment, \$	Percent of subtotal direct investment
Limestone receiving and storage (hoppers, feeders, conveyors, elevators, and bins)	443,000	2.9
Feed preparation (feeders, crushers, elevators, ball mills, benzoic acid storage, conveyor, tanks, and pumps)	572,000	3.8
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, and pumps)	3,203,000	21.4
Sulfur dioxide scrubbers and ducts (4 scrubbers including effluent hold tanks, agitators, pumps, exhaust gas ducts to inlet of fan and mist eliminators with closed-loop wash systems and trap-out trays)	4,170,000	27.8
Stack gas reheat (4 indirect steam reheaters)	556,000	3.7
Fans (4 fans including exhaust gas ducts and dampers between fan and stack gas plenum)	854,000	5.7
Calcium solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, pond, liner, and pond water return pumps)	3,772,000	25.2
Utilities (instrument air generation and supply system, plus distribution systems for obtaining process steam, water, and electricity from the power plant)	67,000	0.4
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	638,000	4.3
Construction facilities	714,000	4.8
Subtotal direct investment	14,989,000	100.0
Engineering design and supervision	1,349,000	9.0
Construction field expense	1,649,000	11.0
Contractor fees	749,000	5.0
Contingency	1,492,000	10.0
Subtotal fixed investment	20,235,000	135.0
Allowance for startup and modifications	1,619,000	10.8
Interest during construction (8 1/2% annual rate)	1,619,000	10.8
Total capital investment	23,473,000	156.6

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by indirect steam reheat.
Disposal pond located 1 mile from power plant.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 30. LIMESTONE - BENZOIC ACID SLURRY PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Limestone	160.1 M tons	4.00/t	640,400	8.61
Benzoic acid	410.0 tons	600.00/ton	246,000	3.31
Subtotal raw material			886,400	11.92
Conversion costs				
Operating labor and supervision	26,280 man-hr	8.00/man-hr	210,200	2.82
Utilities				
Steam	492,800 M lb	0.70/M lb	345,000	4.64
Process water	260,000 M gal	0.08/M gal	20,800	0.28
Electricity	73,770 M kWh	0.010/kWh	707,700	9.51
Maintenance				
Labor and material, .08 x 14,969,000			1,199,100	16.12
Analyses			45,600	0.61
Subtotal conversion costs			2,528,400	33.98
Subtotal direct costs			3,414,800	45.90
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			3,497,500	47.02
Overhead				
Plant, 10% of conversion costs			505,700	6.80
Administrative, 10% of operating labor			21,000	0.28
Subtotal indirect costs			4,024,200	54.10
Total annual operating costs			7,439,000	100.00
	Dollars/ton coal burned	Mills/kWh	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	5.67	2.13	25.62	207.33

^a Basis:

Preliminary estimate from recent TVA in house evaluation.
Remaining life of power plant, 50 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit onstream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$23,473,000; subtotal direct investment, \$14,969,000.
Working capital, \$581,700.
Investment and operating cost for disposal of fly ash excluded.

Table 31. LIME SLURRY PROCESS
VENTURI - MOBILE BED SCRUBBING SCHEME
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Investment, \$	Percent of subtotal direct investment
Lime receiving and storage (bins, feeders, conveyors, and elevators)	795,000	5.3
Feed preparation (conveyors, slakers, tanks, agitators, and pumps)	387,000	2.6
Particulate - sulfur dioxide scrubbers and inlet ducts (4 scrubbers including common feed plenums, effluent hold tanks, agitators, and pumps)	3,203,000	21.5
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, effluent hold tanks, agitators, pumps, and exhaust gas ducts to inlet of fans)	4,523,000	30.3
Stack gas reheat (4 indirect steam reheaters)	542,000	3.6
Fans (4 fans including exhaust gas ducts and dampers between fan and stack gas plenum)	695,000	4.7
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps, pond, liner, and pond water return pumps)	3,445,000	23.1
Utilities (instrument air generation and supply system, plus distribution systems for obtaining process steam, water, and electricity from the power plant)	67,000	0.4
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	552,000	3.7
Construction facilities	710,000	4.8
Subtotal direct investment	14,919,000	100.0
Engineering design and supervision	1,343,000	9.0
Construction field expense	1,641,000	11.0
Contractor fees	746,000	5.0
Contingency	1,492,000	10.0
Subtotal fixed investment	20,141,000	135.0
Allowance for startup and modifications	1,611,000	10.8
Interest during construction (8%/annum rate)	1,611,000	10.8
Total capital investment	23,363,000	156.6

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by indirect steam reheat.
Disposal pond located 1 mile from power plant.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 32. LIME SLURRY PROCESS
VENTURI - MOBILE BED SCRUBBING SCHEME
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Lime	81.2 M tons	22.00/ton	1,786,400	21.95
Subtotal raw material			1,786,400	21.95
Conversion costs				
Operating labor and supervision	22,520 man-hr	8.00/man-hr	178,600	2.19
Utilities				
Steam	490,000 M lb	0.70/M lb	343,000	4.21
Process water	241,900 M gal	0.05/M gal	19,400	0.24
Electricity	61,280,000 kWh	0.010/kWh	612,800	7.52
Maintenance				
Labor and material, .08 x 14,919,000			1,193,500	14.65
Analyses			56,500	.69
Subtotal conversion costs			2,555,000	29.26
Subtotal direct costs			4,170,200	51.19
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			3,481,100	42.74
Overhead				
Plant, 20% of conversion costs			476,800	5.85
Administrative, 10% of operating labor			17,000	0.22
Subtotal indirect costs			3,975,000	48.81
Total annual operating costs			8,146,000	100.00
	Dollars/ton coal burned	Mills/kWh	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	6.21	2.55	25.86	227.05

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Remaining life of power plant, 30 yr.
Coal burned, 1,512,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$25,563,000; subtotal direct investment, \$14,919,000.
Working capital, \$751,000.
Investment and operating cost for disposal of fly ash excluded.

Table 33. LIME SLURRY PROCESS
MOBILE BED SCRUBBING SCHEME
SUMMARY OF ESTIMATED FIXED INVESTMENT^B

(500-MW existing coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal; particulates
removed by existing electrostatic precipitator)

	Investment, \$	Percent of subtotal direct investment
Lime receiving and storage (bins, feeders, conveyors, and elevators)	876,000	7.0
Feed preparation (conveyors, slakers, tanks, agitators, and pumps)	436,000	3.5
Sulfur dioxide scrubbers and ducts (4 scrubbers including common feed plenum, mist eliminators, effluent hold tanks, agitators, pumps, and all duct- work between outlet of supplemental fan and stack gas plenum)	4,964,000	39.9
Stack gas reheat (4 direct oil-fired reheaters)	305,000	2.5
Fans (4 fans including ducts and dampers between tie- in to existing duct and inlet to supplemental fan)	1,118,000	9.0
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps, pond, liner, and pond water return pumps)	3,152,000	25.4
Utilities (instrument air generation and supply system, plus distribution systems for obtaining process steam, water, and electricity from the power plant)	335,000	2.7
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	649,000	5.2
Construction facilities	592,000	4.8
Subtotal direct investment	12,427,000	100.0
Engineering design and supervision	1,243,000	10.0
Construction field expense	1,616,000	13.0
Contractor fees	870,000	7.0
Contingency	1,467,000	11.0
Subtotal fixed investment	17,525,000	141.0
Allowance for startup and modifications	1,402,000	11.3
Interest during construction (8%/annum rate)	1,402,000	11.3
Total capital investment	20,527,000	163.6

^B Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by direct oil-fired reheat.
Disposal plnd located 1 mile from power plant.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average
cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 34. LIME SLURRY PROCESS
MOBILE BED SCRUBBING SCHEME
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW existing coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal;
particulate removed by existing electrostatic precipitator)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw material				
Lime	85.0 M tons	22.00/ton	1,850,000	22.08
Subtotal raw material			1,850,000	22.08
Conversion costs				
Operating labor and supervision	22,320 man-hr	8.00/man-hr	178,600	2.16
Utilities				
Fuel oil (No. 6)	4,236,000 gal	0.23/gal	974,300	11.78
Process water	247,300 M gal	0.08/M gal	19,800	0.24
Electricity	56,056,000 kWh	0.010/kWh	560,500	6.78
Maintenance				
Labor and material, .08 x 12,427,000			994,200	12.02
Analyses			56,500	0.44
Subtotal conversion costs			2,765,900	33.42
Subtotal direct costs			4,589,900	55.50
<u>Indirect Costs</u>				
Average capital charges at 15.3% of total capital investment			5,110,000	37.60
Overhead				
Plant, 20% of conversion costs			552,800	6.68
Administrative, 10% of operating labor			17,900	0.22
Subtotal indirect costs			5,680,700	44.50
Total annual operating costs			8,270,600	100.00
	Dollars/ton coal burned	Mills/kWh	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	6.16	2.36	25.69	225.48

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Remaining life of power plant, 25 yr.
Coal burned, 1,341,700 tons/yr, 9,200 Btu/kWh.
Stack gas reheat to 175°F.
Power unit onstream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$20,527,000; subtotal direct investment, \$12,427,000.
Working capital, \$500,000.
Investment and operating cost for removal and disposal of fly ash excluded.

Table 35. LIME SLURRY PROCESS WITH ONSITE LIME PRODUCTION
VENTURI - VENTURI SCRUBBING SCHEME
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	<u>Investment, \$</u>	<u>Percent of subtotal direct investment</u>
Limestone receiving and storage (hoppers, feeders, conveyors, elevators, and bins)	751,000	4.3
Limestone calcination (rotary kiln, fan, conveyors, elevators, dust collector, and coal storage)	3,041,000	17.3
Feed preparation (conveyors, slakers, tanks, agitators, and pumps)	387,000	2.2
Particulate - sulfur dioxide scrubbers and inlet ducts (4 scrubbers including common feed plenum and pumps)	4,017,000	22.9
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, pumps, and exhaust gas ducts to inlet of fans)	3,153,000	18.0
Stack gas reheat (4 indirect steam reheaters)	542,000	3.1
Fans (4 fans including exhaust gas ducts and dampers between fan and stack gas plenum)	767,000	4.4
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps, pond, liner, and pond water return pumps)	3,356,000	19.1
Utilities (instrument air generation and supply system, plus distribution systems for obtaining process steam, water, and electricity from the power plant)	74,000	0.4
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	607,000	3.5
Construction facilities	845,000	4.8
Subtotal direct investment	17,530,000	100.0
Engineering design and supervision	1,578,000	9.0
Construction field expense	1,908,000	11.0
Contractor fees	877,000	5.0
Contingency	1,753,000	10.0
Subtotal fixed investment	23,664,000	135.0
Allowance for startup and modifications	1,893,000	10.8
Interest during construction (8%/annum rate)	1,893,000	10.8
Total capital investment	27,452,000	156.6

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by indirect steam reheat.
Disposal pond located 1 mile from power plant.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 36. SODIUM SOLUTION - H_2SO_4 PRODUCTION PROCESS
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO_2 removal; 13.3 tons/hr 100% H_2SO_4)

	Investment, \$	Percent of subtotal direct investment
Soda ash and antioxidant receiving, storage, and preparation (pneumatic conveyor and blower, feeders, mixing tank, agitator, and pumps)	225,000	1.2
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, pumps, and fly ash neutralization facilities)	3,846,000	20.6
Sulfur dioxide scrubbers and ducts (4 scrubbers including mist eliminators, pumps, and exhaust gas ducts to inlet of fan)	4,269,000	22.9
Stack gas reheat (4 indirect steam reheaters)	539,000	2.9
Fans (4 fans including exhaust gas ducts and dampers between fans and stack gas plenum)	899,000	4.8
Purge treatment (refrigeration system, chiller-crystallizer, feed coolers, centrifuge, rotary dryer, steam/air heater, fan, dust collectors, feeders, tanks, agitators, pumps, conveyors, elevator, and bins)	1,473,000	7.9
Sulfur dioxide regeneration (evaporator-crystallizers, heaters, condensers, strippers, desuperheater, tanks, agitators, and pumps)	2,622,000	14.1
Sulfuric acid plant (complete contact unit for sulfuric acid production)	2,793,000	15.0
Sulfuric acid storage (storage and shipping facilities for 30 days production of H_2SO_4)	242,000	1.3
Utilities (instrument air generation and supply system, and distribution systems for obtaining process steam, water, and electricity from power plant)	195,000	1.0
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	662,000	3.5
Construction facilities	898,000	4.8
Subtotal direct investment	18,643,000	100.0
Engineering design and supervision	2,051,000	11.0
Construction field expense	2,051,000	11.0
Contractor fees	932,000	5.0
Contingency	1,864,000	10.0
Subtotal fixed investment	25,541,000	137.0
Allowance for startup and modifications	2,554,000	13.7
Interest during construction (8% annum rate)	2,042,000	11.0
Total capital investment	30,138,000	161.7

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by indirect steam reheat.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
Fly ash slurry neutralized before disposal; closed loop water utilization for first stage.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 37. SODIUM SOLUTION - H_2SO_4 PRODUCTION PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO_2 removal; 92,000 tons/yr 100% H_2SO_4)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw materials				
Lime (1st stage neutralization)	154.1 tons	26.00/ton	3,500	0.03
Soda ash	9,500 tons	52.00/ton	483,600	4.17
Antioxidant	317,100 lb	2.00/lb	634,200	5.47
Catalyst	1,800 liter	1.65/liter	2,970	0.10
Subtotal raw materials			1,124,300	9.77
Conversion costs				
Operating labor and supervision	45,800 man-hr	8.00/man-hr	366,900	3.21
Utilities				
Steam	2,157,800 M lb	0.70/M lb	1,496,500	12.90
Process water	11,672,400 M gal	0.02/M gal	233,400	1.72
Electricity	79,534,000 kWh	0.010/kWh	795,300	6.39
Maintenance				
Labor and material, .06 x 18,643,000			1,118,600	9.75
Analyses			108,000	0.95
Subtotal conversion costs			4,118,700	38.98
Subtotal direct costs			5,243,000	48.75
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			4,490,600	39.16
Overhead				
Plant, 20% of conversion costs			823,700	7.80
Administrative and marketing			497,500	4.29
Subtotal indirect costs			5,811,600	51.25
Total annual operating cost			11,054,600	100.00
	Dollars/ton 100% H_2SO_4	Dollars/ton coal burned	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	119.01	8.42	35.09	306.57

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Remaining life of power plant, 30 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit onstream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$30,138,000; subtotal direct investment, \$18,643,000.
Working capital, \$948,500.
Investment and operating cost for disposal of fly ash excluded.

Table 38. AMMONIA LIMESTONE/LIME DOUBLE ALKALI PROCESS
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(900-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	Investment, \$	Percent of subtotal direct investment
Ammonia, limestone, and lime receiving and storage (bins, feeders, conveyors, elevators, and ammonia storage sphere)	937,000	4.7
Feed preparation (feeders, crushers, elevators, ball mills, vaporizer, conveyors, slaker, tanks, and pumps)	681,000	3.4
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, and pumps)	3,846,000	19.4
Sulfur dioxide scrubbers and ducts (4 scrubbers including wire mesh mist eliminators, effluent hold tanks, pumps, and exhaust gas ducts to inlet of fan)	4,219,000	21.3
Stack gas reheat (4 indirect steam reheaters)	539,000	2.7
Fans (4 fans including exhaust gas ducts and dampers between fans and stack gas plenum)	889,000	4.5
Regeneration and precipitates handling (thickeners, filters, tanks, agitators, conveyors, and pumps)	3,581,000	18.0
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps and pond)	3,508,000	17.7
Utilities (instrument air generation and supply system, and distribution systems for obtaining process steam, water, and electricity from the power plant)	67,000	0.3
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	638,000	3.2
Construction facilities	945,000	4.8
Subtotal direct investment	19,850,000	100.0
Engineering design and supervision	1,787,000	9.0
Construction field expense	2,184,000	11.0
Contractor fees	993,000	5.0
Contingency	1,095,000	10.0
Subtotal fixed investment	26,799,000	135.0
Allowance for startup and modifications	2,144,000	10.8
Interest during construction (7% annual rate)	2,144,000	10.8
Total capital investment	31,087,000	156.6

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by indirect steam reheat.
Midwest plant location represents project beginning mid-1972, ending mid-1975. Average
cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
System is designed for closed loop water utilization.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 39. AMMONIA LIMESTONE/LIME DOUBLE ALKALI PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel; 90% SO₂ removal)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw materials				
Limestone	142.0 M tons	4.00/ton	568,000	5.87
Lime	15.3 tons	22.00/ton	336,600	3.48
Ammonia	2,110 tons	130.00/ton	274,300	2.84
Subtotal raw materials			1,178,900	12.19
Conversion costs				
Operating labor and supervision	35,100 man-hr	8.00/man-hr	280,800	2.90
Utilities				
Steam	883,400 M lb	0.70/M lb	618,400	6.40
Process water	262,000 M gal	0.08/M gal	21,000	0.22
Electricity	63,200,000 KWh	0.010/kWh	632,000	6.54
Maintenance				
Labor and material, .08 x 19,850,000			1,588,000	16.42
Analyses			51,700	0.53
Subtotal conversion costs			3,191,900	33.01
Subtotal direct costs			4,370,800	45.20
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			4,632,000	47.91
Overhead				
Plant, 20% of conversion costs			638,400	6.60
Administrative, 10% of operating labor			28,100	0.29
Subtotal indirect costs			5,298,500	54.80
Total annual operating costs			9,669,300	100.00
	Dollars/ton coal burned	Mills/kWh	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	7.37	2.76	30.70	269.49

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Remaining life of power plant, 30 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$31,087,000; subtotal direct investment, \$19,850,000.
Working capital, \$746,100.
Investment and operating cost for disposal of fly ash excluded.

Table 40. SODIUM Limestone DOUBLE ALKALI PROCESS
SUMMARY OF ESTIMATED FIXED INVESTMENT^a

(500-MW new coal-fired power unit, 3.5% S in fuel;
90% SO₂ removal; onsite solids disposal)

	<u>Investment, \$</u>	<u>Percent of subtotal direct investment</u>
Limestone, soda ash, and antioxidant receiving and storage (bins, hoppers, feeders, conveyors, elevators, and pumps)	319,000	1.6
Feed preparation (feeders, weigh belts, agitators, crushers, ball mills, tanks, and pumps)	607,000	3.1
Particulate scrubbers and inlet ducts (4 scrubbers including common feed plenum, effluent hold tanks, agitators, and pumps)	3,846,000	19.6
Sulfur dioxide scrubbers and ducts (4 scrubbers including wire mesh mist eliminators, effluent hold tanks, pumps, and exhaust gas ducts to inlet of fan)	4,364,000	22.2
Stack gas reheat (4 indirect steam reheaters)	539,000	2.7
Fans (4 fans including exhaust gas ducts and dampers between fans and stack gas plenum)	1,170,000	6.0
Regeneration and precipitates handling (sulfate reactor absorber and thickener, filter, conveyor, steam coil, tank, agitator, and pumps)	3,583,000	18.2
Calcium solids disposal (onsite disposal facilities including slurry disposal pumps and pond)	3,551,000	18.1
Utilities (instrument air generation and supply system, and distribution systems for obtaining process steam, water, and electricity from power plant)	72,000	0.4
Service facilities (buildings, shops, stores, site development, roads, railroads, and walkways)	656,000	3.3
Construction facilities	92,000	0.5
Subtotal direct investment	19,142,000	100.0
Engineering design and supervision	1,708,000	9.0
Construction field expense	2,161,000	11.0
Contractor fees	980,000	5.0
Contingency	1,664,000	10.0
Subtotal fixed investment	26,517,000	135.0
Allowance for startup and modifications	3,121,000	10.8
Interest during construction (8%/annum rate)	2,121,000	10.8
Total capital investment	30,759,000	156.6

^a Basis:

Preliminary estimate from recent TVA in-house evaluation.
Stack gas reheat to 175°F by indirect steam reheat.
Midwest plant location represents project beginning mid-1973, ending mid-1975. Average cost basis for scaling, mid-1974.
Minimum in process storage; only pumps are spared.
System is designed for closed loop water utilization.
Investment requirements for disposal of fly ash excluded.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 41. SODIUM LIMESTONE DOUBLE ALKALI PROCESS
TOTAL AVERAGE ANNUAL OPERATING COSTS - REGULATED UTILITY ECONOMICS^a

(500-MW new coal-fired power unit, 3.5% S in fuel; 90% SO₂ removal)

	Annual quantity	Unit cost, \$	Total annual cost, \$	Percent of total annual operating cost
<u>Direct Costs</u>				
Delivered raw materials				
Limestone	177.5 M tons	4.00/ton	710,000	7.44
Soda ash	834.0 tons	52.00/ton	43,400	0.45
Antioxidant	240.0 M lb	2.00/lb	480,000	5.03
Subtotal raw materials			1,233,400	12.92
Conversion costs				
Operating labor and supervision	30,600 man-hr	8.00/man-hr	260,800	2.73
Utilities				
Steam	508,400 M lb	0.70/M lb	355,900	3.73
Process water	260,000 M gal	0.08/M gal	20,800	0.22
Electricity	82,840,000 kWh	0.010/kWh	828,400	8.68
Maintenance				
Labor and material, .08 x 19,642,000			1,571,400	16.46
Analyses			48,700	0.51
Subtotal conversion costs			3,086,000	32.33
Subtotal direct costs			4,319,400	45.25
<u>Indirect Costs</u>				
Average capital charges at 14.9% of total capital investment			4,583,100	48.01
Overhead				
Plant, 20% of conversion costs			617,200	6.47
Administrative, 10% operating labor			26,100	0.27
Subtotal indirect costs			5,226,400	54.75
Total annual operating cost			9,545,800	100.00
	Dollars/ton coal burned	Mills/kWh	Cents/million Btu heat input	Dollars/ton sulfur removed
Equivalent unit operating cost	7.27	2.73	30.30	266.05

^a

Basis:

Preliminary estimate from recent TVA in-house evaluation.
Remaining life of power plant, 30 yr.
Coal burned, 1,312,500 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Power unit on-stream time, 7,000 hr/yr.
Midwest plant location, 1975 operating costs.
Total capital investment, \$30,759,000; subtotal direct investment, \$19,642,000.
Working capital, \$739,200.
Investment and operating cost for disposal of fly ash excluded.

3. The Cat-Ox process has the poorest investment economy of scale (unit size) of all five systems; however, for sulfur content of fuel, Cat-Ox has the best scale factor. A reheat (existing) Cat-Ox system requiring full particulate removal facilities to .005 gr/scf was found to be only 2.5% higher than a new integrated unit.
4. Plant age is an important factor only in the limestone and lime processes where pond size depends on remaining plant life; older units can best use limestone or lime processes.
5. Removal of only 80% of the SO_2 (which would meet Federal new source emission standards for 3.5% S) instead of 90% decreases investment by only 3 to 5%.

Operating Cost

1. For both 3.5% S coal-fired and 2.5% S oil-fired power units under the premises used, limestone has the lowest annual operating cost and sodium the highest. Even after credits for recovery products are applied, the lifetime operating costs are higher for the regenerable systems than for the throwaway systems. Cat-Ox becomes competitive with limestone on new oil-fired units, especially for smaller units.
2. As sulfur content of fuel increases, the relative operating cost for all five systems changes. Cat-Ox lifetime operating costs go from highest at low sulfur levels to third on coal-fired units and first on oil-fired units with high-sulfur fuel levels. These are some of the most interesting results of the study. The heat credit for Cat-Ox becomes significant at high sulfur levels.
3. The lime and sodium processes rank high in raw material cost and magnesia and Cat-Ox are low. Limestone has the largest range in cost of all the raw materials.
4. Sodium scrubbing has the highest total labor cost and Cat-Ox the lowest; however, labor is one of the smallest components (1-3% of total) for all five processes.
5. Energy costs are significant for all systems; applications of sodium scrubbing on existing units require the greatest amount of energy (35% of total operating cost). If a double-effect evaporator-crystallizer is utilized in the sodium system, overall energy costs could be reduced as much as 18%. The magnesia process is also energy intensive; the Cat-Ox system uses the lowest amount (5%).
6. Although expense for antioxidant (sodium system) is high, it is justified to keep sulfate formation down and save Na_2CO_3 makeup.

7. Maintenance is quite significant ranging from 7% of total operating cost for a Cat-Ox case to 17% for a limestone case.
8. Capital charges are the largest individual component of operating cost for all five processes. For new Cat-Ox systems capital costs are 72% of total. For the other processes, capital charges are in the range of 39 to 50%. A change in depreciation rate or cost of money will obviously affect Cat-Ox the most.
9. Only about 4 to 5% of total operating cost is saved when 80% SO₂ removal is provided instead of 90%.
10. A Cat-Ox process on an existing plant has a 40% greater operating cost than on a new system. This is caused by the high energy required for stack gas reheat from 310 to 890°F prior to conversion of SO₂ to SO₃.
11. As would be expected, the scrubbing steps of each process are the highest cost operations.
12. For the energy intensive processes (magnesia, sodium), oil-fired systems are at a disadvantage because of the high cost of fuel oil relative to coal (\$1.53/MM Btu compared to \$0.54/MM Btu).
13. Sale of byproducts at the values assumed in the study would reduce the base case lifetime operating cost 7.2% for the magnesia, 7.1% for the sodium, and 4.9% for the Cat-Ox process.
14. Because of product revenues, the relative ranking of the magnesia scrubbing process on oil-fired units improves under lifetime operating costs until it is the lowest cost system above 800-MW size.
15. Labor cost escalation (7.5% per year) over the project life would add about 7-10% to total project cost.
16. Regardless of which process is utilized, the increase in the cost of power to consumers for the base case is projected to range from 2.86 to 3.80 mills/kWh. For all case variations, projected costs could range from 1.57 to 7.90 mills/kWh.

Although there are differences in the depth and quality of the EPA-sponsored evaluations and the TVA in-house estimates, some additional conclusions can be derived from comparing all the base case results as shown in Table 42.

1. For limestone scrubbing, process savings from the addition of benzoic acid to reduce limestone stoichiometry (and therefore raw material and solids disposal costs) are relatively small.

Table 42. COMPARISON OF RESULTS OF EPA-SPONSORED EVALUATIONS AND TVA IN-HOUSE ESTIMATES
NEW 500-MW, 3.5% SULFUR COAL-FIRED UNITS, 30 YEARS LIFE (EXCEPTIONS NOTED)

Process	Results		
	EPA-sponsored evaluations	TVA in-house estimates	
	Capital investment, \$	Average annual operating cost, \$	Capital investment, \$ Average annual operating cost, \$
Limestone			
Limestone slurry process	25,163,000	7,702,700	- 7,439,000
Limestone-benzoic acid slurry process	-	-	23,473,000
Lime			
Lime slurry process (venturi-venturi scheme)	22,422,000	9,101,900	23,363,000 9,146,000
Lime slurry process (venturi-mobile bed scheme)	-	-	-
Lime slurry process (venturi-venturi scheme) ^a	26,027,000 ^a	9,612,400 ^a	20,327,000 ^a 9,270,600 ^a
Lime slurry process (mobile bed scheme) ^a	-	-	27,452,000 8,101,900
Lime slurry process with onsite lime production (venturi-venturi scheme)	-	-	-
Magnesia			
Magnesia slurry - regeneration process	26,406,000	9,210,900	- -
Sodium			
Sodium solution - SO ₂ reduction process	30,491,000	11,601,500	30,138,000 11,054,800
Sodium solution - MgSO ₄ production process	-	-	-
Cat-Ox			
Catalytic oxidation process	42,736,000	9,873,900	- -
Double alkali			
Ammonia-limestone/lime double-alkali process	-	-	31,087,000 9,669,300
Sodium-limestone double-alkali process	-	-	30,759,000 9,545,900

^a Existing 500-MW, 3.5% sulfur coal-fired unit, 25 years remaining life.

2. Costs for a two-stage venturi lime system and a venturi - mobile bed lime system on a new coal-fired unit are reasonably close; however, for an existing coal-fired power plant or oil-fired unit not requiring additional particulate removal facilities, the mobile bed scrubber option requires about 22% less capital and 14% less operating cost.
3. Production of sulfuric acid is slightly less expensive than production of sulfur when sodium scrubbing - regeneration is used. Depending on prevailing market value of byproduct acid and sulfur, the expected revenues could determine which option is best (2.8 times as much acid as sulfur).
4. Based on the "state of the art" at this time, the costs of the two double-alkali options are about the same, but at least 20-25% more costly than lime or limestone scrubbing.

EPA/RTP PILOT STUDIES RELATED TO UNSATURATED
OPERATION OF LIME AND LIMESTONE SCRUBBERS

by

Robert H. Borgwardt
Control Systems Laboratory
U.S. Environmental Protection Agency
National Environmental Research Center
Research Triangle Park, North Carolina

ABSTRACT

An oddity of lime and limestone SO_2 scrubbers is the observation that they can run unsaturated with respect to dissolved $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ during closed loop operation. This fact offers a unique opportunity, if the phenomenon responsible can be understood and controlled, to effectively reduce scaling potential which has persisted as one of the principal reliability problems of those systems. In addition to eliminating the potential for scaling of absorber internals, unsaturated operation could greatly increase the amount of clarified liquor available for mist eliminator washing. Experimental studies of the process conditions necessary to achieve unsaturated scrubbing liquor are reported, and a mechanism is proposed that accounts for it. It is shown that the sulfate formed by oxidation is purged, without crystallization, as a "solid solution" within the precipitated CaSO_3 . The amount of sulfate that can be thus incorporated is related to the sulfate activity in the scrubbing liquor and the SO_2 precipitation rate in the hold tank. The effect of Cl^- on suppressing sulfate activity and the effect of Mg^{++} on raising sulfate activity are evaluated and correlated with oxidation and saturation. The observed superiority of lime vs. limestone with respect to ease of development of unsaturated conditions is attributed to the higher precipitation rates and low oxidation that are characteristic of lime systems.

EPA/RTP PILOT STUDIES RELATED TO UNSATURATED
OPERATION OF LIME AND LIMESTONE SCRUBBERS

SATURATED VS. UNSATURATED MODES

Calcium sulfate is generated in the scrubbing liquor when dissolved SO_2 is oxidized either in the scrubbing tower, where the liquor is exposed to flue gas containing about 5% oxygen, or in the ancillary process vessels, where it is exposed to the air. Unlike calcium sulfite, the sulfate is not precipitated from solution by the rise in pH that occurs in the hold tank and, consequently, accumulates in the scrubbing liquor until it is supersaturated with $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$. Calcium sulfate crystallizes at a slow rate and its accumulation in the liquor of a closed-loop scrubbing system ceases only when a level of supersaturation is reached that yields a crystallization rate equal to the total oxidation rate in all parts of the system - a necessary condition for steady state operation. Present methods for reducing the steady state supersaturation thus follow the crystallization approach by 1) maximizing the rate of crystallization with high concentrations of seed crystals, and 2) providing maximum residence time for the slurry in the hold tank (crystallizer) before returning it to the scrubber. Together, these factors can reduce the level of supersaturation of the scrubber feed liquor sufficiently to prevent scale growth. These principles were successfully demonstrated by I.C.I. at 1000 ppm SO_2 scrubber feed concentration, and present practice is based largely on their experience.

The possibility of achieving saturation levels close to, or even below the equilibrium solubility of $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ in a lime scrubber liquor was first indicated by J. Martin of Combustion Engineering, Inc. in discussing the Louisville G&E operations at the 3rd EPA Symposium in May 1973. Since subsaturated liquor is contradictory to the accepted crystallization mechanism for sulfate purge, the reported unsaturated operation was initially regarded as evidence of either open loop conditions or total lack of oxidation. It is interesting to note that the I.C.I. scrubber data reported in 1935, in which tests of supersaturation were conducted on a regular basis, showed several periods of negative supersaturation. No comment was made on the anomaly although it sometimes persisted for several days. Inspection of the I.C.I. reports reveals that these values were generally associated with periods of low oxidation; 12-20%. It is also significant that a relatively short hold-tank residence time was used in that system.

In addition to the I.C.I. and L.G.& E. scrubbers, Mitsui appears to be another full-size unit that has achieved unsaturated conditions. The absence of chloride in the scrubbing liquor of each of these systems is an important factor they all have in common that, as will be shown, has no doubt contributed to their success. The use of lime feed in the latter two cases is another factor.

The capability of operating a limestone scrubber at steady-state unsaturated conditions, in which loop closure could be proven by material balance, was demonstrated in the EPA/RTP model in May 1973. These tests involved a plug flow hold tank in which no air contacted the circulating slurry, and oxidation values were only about 10%. Since no chloride was present, unsaturated liquor was clearly evidenced by calcium concentrations of only 100-200 ppm, less than 1/3 the saturated concentration. Material balances confirmed loop closure at 62% solids and showed no unaccounted-for liquor loss. Since that time, extensive experimentation has been carried out at RTP with both limestone and lime scrubbers to define the process conditions required to maintain operation in the unsaturated mode. This report summarizes that work.

The most important potential advantage of the unsaturated mode of operation is that it would allow virtually unlimited washing of the mist eliminators - the critical component in the overall reliability of most systems at present - with liquor that would not contribute to chemical scaling of the mist eliminator surfaces.

MECHANISM OF SULFATE PURGE IN THE UNSATURATED MODE

Proof of unsaturated operation in a closed loop scrubber requires a mechanism of sulfate precipitation from the liquid phase that does not involve crystallization. Given the fact that the liquor is unsaturated, it follows that crystallization of calcium sulfate will not occur and cannot account for the presence of sulfate in the solid. It is nevertheless observed that significant amounts of sulfate are present in the solids when a scrubber is running unsaturated. Furthermore, the incorporation of sulfate within the solid is a necessary condition for steady-state unsaturated operation since the sulfate generated by oxidation cannot be purged in sufficient amounts any other way. For example, it can be shown that the maximum amount of dissolved sulfate lost in the liquid purge from a closed loop scrubber (with filter cake moisture) is equivalent to only about 0.2% oxidation when the chloride level is 5000 ppm.

A mechanism whereby sulfate can be incorporated by calcium sulfite without crystallization was proposed by I.C.I. in 1951 to account for the unusual properties of the solid product from their Bankside scrubber. Examination of those solids, which chemical analysis showed to contain about 10% calcium sulfate, revealed no gypsum or anhydrite detectable by x-ray diffraction. It was postulated from this result and from measurements of the degree of hydration of the calcium sulfate that it was present as a homogeneous "solid solution" within the calcium sulfite. Laboratory experiments conducted at simulated scrubber conditions confirmed that sulfate was brought down with the precipitating CaSO_3 in amounts equivalent to a maximum ratio of 0.23 mole sulfate per mole sulfite.

Tests similar to the above were carried out on solids obtained from the EPA/RTP scrubber during unsaturated operation. These solids, which contained 10-15 mole percent sulfate by chemical analysis, showed no calcium sulfate when examined by x-ray diffraction in several different laboratories. Stirring the solids with distilled water would not extract the calcium sulfate, confirming that it was not present as a separate phase. Dissolution of the solid with HCl, followed by water extraction and recrystallization, yields pure gypsum in amounts expected from the chemical analysis of the original solid. The characteristics of the solid conform in all respects to those reported by I.C.I. for solid solutions and it was concluded that the formation of this product constitutes the mechanism by which sulfate is purged from scrubbers during unsaturated operation. The I.C.I. terminology has been adopted here when referring to this compound although recent evidence suggests that it is not a true solid solution in the thermodynamic sense.

OXIDATION AND SATURATION

When a scrubber is running unsaturated and the sulfate is being purged entirely in the form of solid solution, a material balance shows that the rate of solid solution formation must equal the overall rate of oxidation. If this condition were not satisfied, the excess sulfate would accumulate in the closed loop system and the liquor would, in time, become saturated regardless of solid solution formation:

$$\text{Input} = \text{Output} + \text{Accumulation (or depletion)} \quad (1)$$

For steady-state unsaturated operation, accumulation = 0 and since sulfate cannot leave the system by any means other than solid solution, equation (1) reduces to:

$$\frac{1}{\text{ratio}} = \frac{1}{\text{oxidation}} - 1 \quad (2)$$

where ratio is moles sulfate precipitated as solid solution per mole of precipitated sulfite.

The relationship expressed by equation (2) is a necessary condition for unsaturated operation at any level of subsaturation. If conditions are such that equation (2) cannot be satisfied, CaSO_4 accumulation will occur and the system must become saturated. Inspection of the equation shows that this will happen if oxidation exceeds the value corresponding to the maximum limit of the value of ratio.

Figure 1 is a plot of scrubber feed saturations observed in the EPA pilot plant over the full range of steady state oxidation values. These data represent chloride-free operation with limestone feed, consequently the calcium concentration is a direct index of saturation, with $\text{Ca} \approx 630$ ppm corresponding to saturated liquor. The data show a continuous increase in saturation from about 1/6x at 10% oxidation to 1x at about 20% oxidation which, according to the discussion above, is equivalent to the maximum mole ratio under the test conditions. The maximum mole ratio at 20% oxidation is thus, according to equation (2), in agreement with the value determined by I.C.I. from lab data.

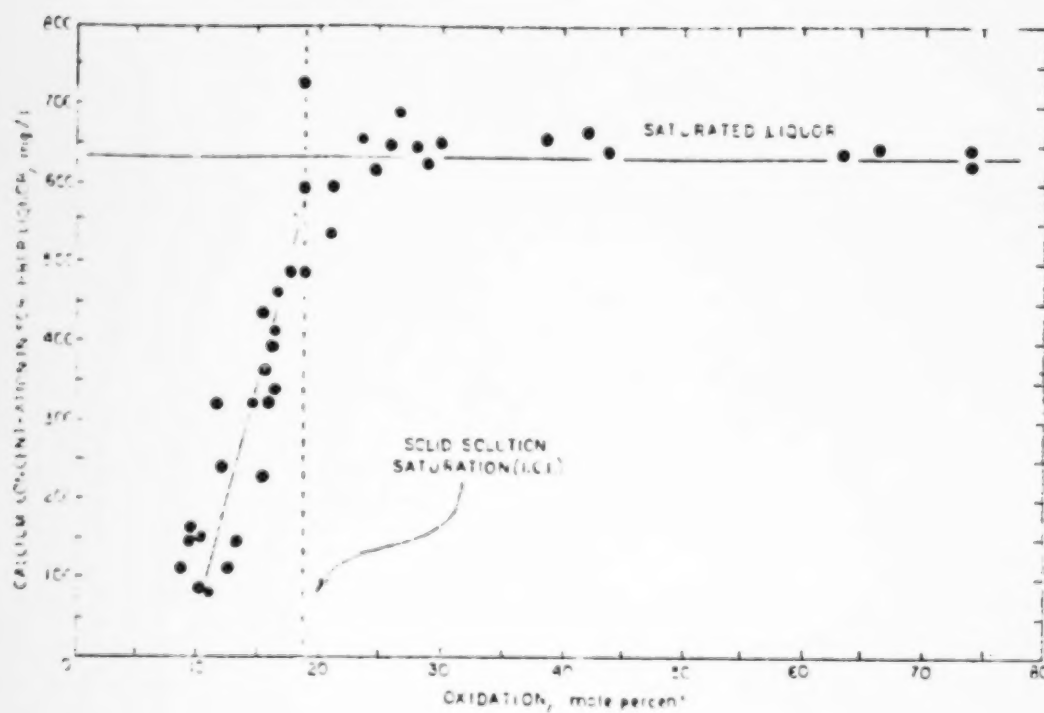


Figure 1. Calcium Sulfate Saturation as a Function of Oxidation
In Limestone Scrubber. H_2 = 300-900 ppm, Cl = 0.

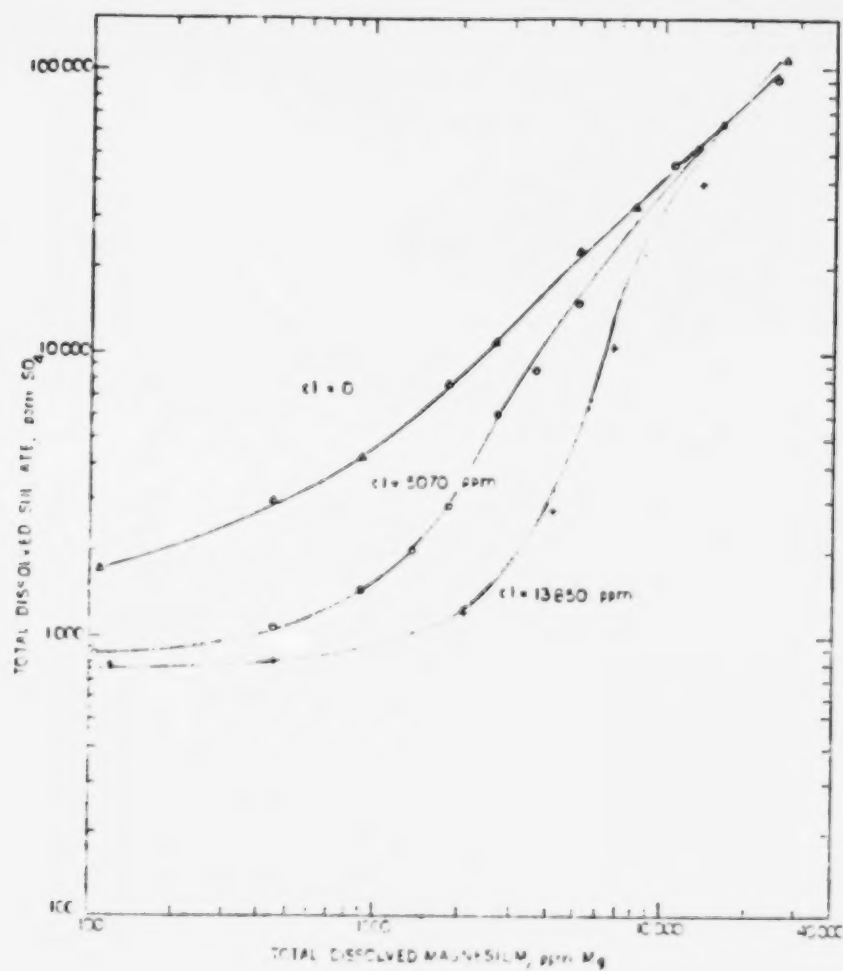


Figure 2. Solubility of $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ in Solutions Containing Chloride and Magnesium. Laboratory Measurements at 25°C.

MAXIMUM MOLE RATIO OF SULFATE TO SULFITE IN SOLID SOLUTION

When sulfite precipitates from a liquor containing dissolved sulfate, it is reasonable to expect that the amount of sulfate incorporated as solid solution is dependent upon the concentration of sulfate in the liquor. As shown by the solubility data of Figure 2, the equilibrium concentration of sulfate in solution can vary over two orders of magnitude, in lime/limestone scrubbers, depending upon the steady state concentrations of Mg^{++} and Cl^- . At a given chloride level, the magnesium concentration in a closed loop scrubber increases with the amount of soluble magnesium fed, accumulating primarily as $MgSO_4^o$ and to a lesser extent, as $CaSO_4^o$ and $SO_4^{=}$.

Experiments using limestone feed, to which varying amounts of Mg were added (as MgO) to build up the concentration of total dissolved sulfates, confirmed that solid-solution mole ratios exceeding 0.23 could be achieved by operation at high Mg^{++} levels. Similarly, it was shown that either $MgSO_4$ or MgO addition to a lime scrubber, which was initially operating saturated, would force the system into the unsaturated mode. From these results it was concluded that the mole ratio of sulfate to sulfite in the solid is a function of the sulfate concentration in the liquor from which it precipitates.

The presence of chloride in scrubber liquors is unavoidable in closed-loop systems operating on coal-fired power plants that burn coal containing chloride. This chloride enters the scrubber with the flue gas as HCl, usually in concentrations less than 100 ppm. It accumulates in the scrubber liquor because, unlike SO_2 and SO_3 , it forms no insoluble calcium compounds and can therefore leave the system only with the liquid purge, i.e. filter cake moisture or clarifier underflow. Since the liquid purge from closed loop scrubbers is small, high steady state levels of chloride must be reached in order for its purge rate to equal the rate of chloride absorption in the tower. At Shawnee, Cl^- is the most abundant dissolved specie in the scrubbing liquor.

Many runs have now been made in the unsaturated mode with both lime and limestone scrubbers, at varying levels of chloride, magnesium and oxidation. In Figure 3 the mole ratios of sulfate/sulfite in the solids are plotted against the sulfate ion activities calculated from the average scrubber feed liquor analyses for those runs. It is apparent from Figure 3 that the value of ratio increases in an approximately linear manner with sulfate activity. This correlation is significant in two respects: 1) an increase in ratio as any function of sulfate activity is sufficient condition to qualitatively explain the formation of unsaturated solutions (given this relationship, the existence of an unsaturated mode is a predictable consequence); and 2) a quantitative relationship between these

two variables will enable one to calculate the degree of subsaturation that will result from any given levels of chloride, magnesium and oxidation. This is accomplished through simultaneous solution of this relationship with six other equations that define ionic strength, ion balance, sulfate activity and saturation. Using a simple power function to represent Figure 3, these relationships are:

$$\text{Ion balance:} \quad \text{Ca} = 40 \left[\frac{\text{SO}_4}{96} + \frac{\text{Cl}}{71} + \frac{\text{SO}_2}{64} - \frac{\text{Mg}}{24} \right] \quad (3)$$

$$\text{Ionic strength:} \quad \mu = 0.5 \left[\frac{\text{SO}_4}{24} + \frac{\text{Ca}}{10} + \frac{\text{Mg}}{6} - \frac{\text{SO}_2}{16} + \frac{\text{Cl}}{35.5} \right] \quad (4)$$

$$\text{SO}_4^= \text{ act. coeff.:} \quad \text{Log } \alpha = \frac{-2.02 \sqrt{\mu}}{1+0.7 \sqrt{\mu}} \quad (5)$$

$$\text{SO}_4^= \text{ activity:} \quad \left\{ \begin{array}{l} \ln c = \frac{-2.227 - \ln \mu}{2.041} \end{array} \right. \quad (6)$$

$$\text{SO}_4^= \text{ act.} = \frac{\alpha}{\sqrt{c}} \left[\frac{\text{SO}_4}{96} \right] \quad (7)$$

$$\text{Solid Soln. formation:} \quad \ln \text{SO}_4^= \text{ act.} = \frac{-2.395 + \ln \text{ratio}}{0.906} \quad (8)$$

$$\text{Saturation:} \quad \text{Sat.} = \frac{\alpha^2}{c} \left[\frac{\text{SO}_4}{96} \right] \left[\frac{\text{Ca}}{40} \right] \frac{1}{3.32 \times 10^{-5}} \quad (9)$$

For any given values of oxidation, Cl, Mg and SO₂ in the scrubbing liquor, the values of the other 8 variables can be determined from equations (2) - (9): Ca, SO₄, μ, C, α, SO₄⁼ activity, ratio, and saturation.

Solution of these equations for a limestone scrubber is shown in Figure 4, where the Mg levels required for 0.8x saturation of the scrubber feed are plotted against oxidation at several levels of chloride.

PRECIPITATION RATE

In addition to the values of oxidation and sulfate activity another constraint must also be satisfied for a system to operate in the unsaturated mode. The data shown in Figures 1 and 3 were all obtained at residence times of about 6 minutes, inlet SO₂ concentrations of 3000 ppm, and scrubber gas velocities of 9 ft/sec or more. It is observed that, when all other variables remain constant, an increase in hold tank residence time to 12 minutes will cause a limestone scrubber to operate supersaturated even though the oxidation, magnesium, and chloride levels are maintained at values that yield unsaturated liquor at the shorter residence time. It is evident that solid solution formation is not simply a function of the liquor composition, but kinetic factors are also involved. Table 1 compares steady state data from duplicate limestone/MgO scrubber runs at different hold tank residence times. The scrubber ran supersaturated for 4 weeks at 12 min. residence time, where-

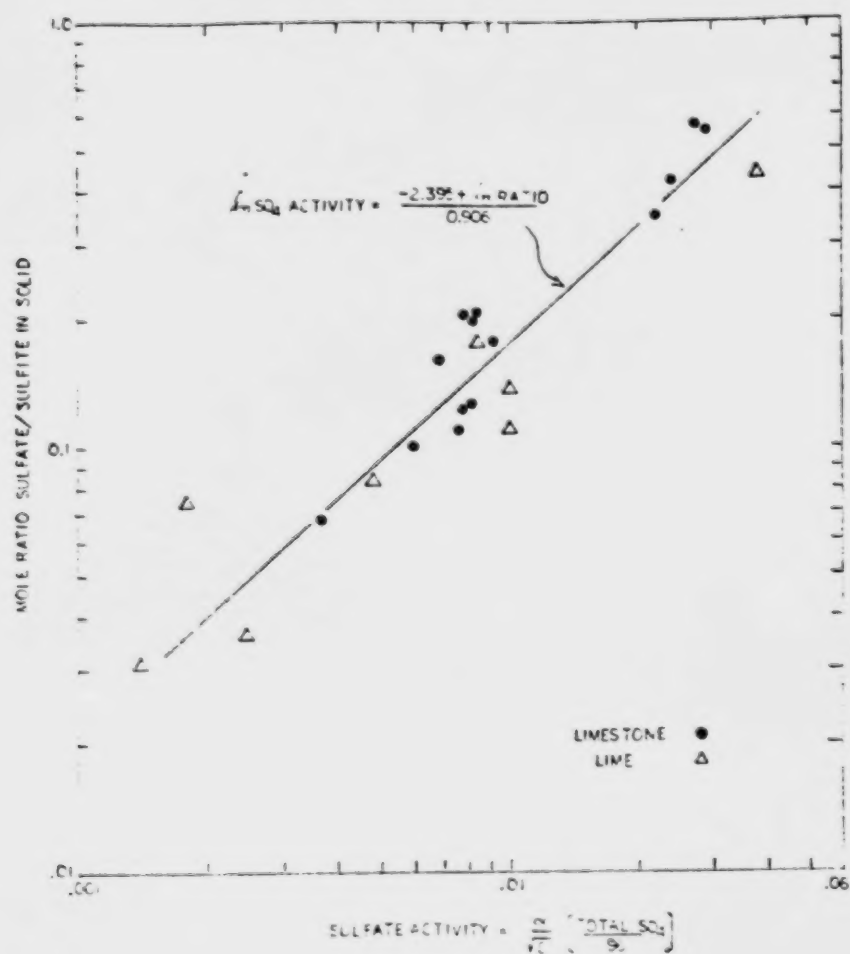


Figure 3. Mole Ratio of Sulfate to Sulfite in Pilot Plant Solids, as a Function of the Sulfate Ion Activity in the Scrubbing Liquor. Saturation < 0.9x

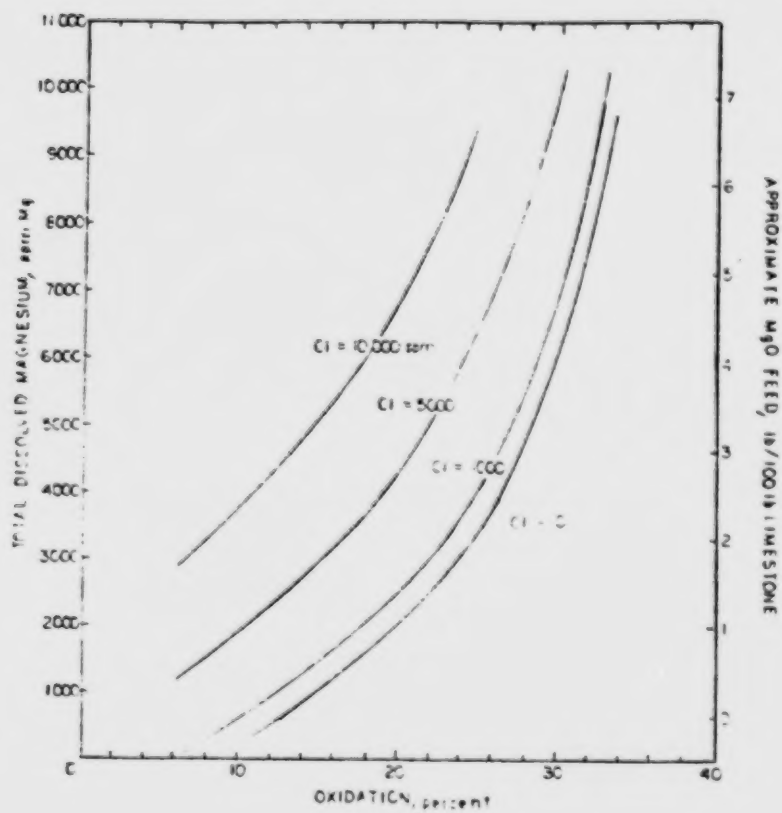


Figure 4. Magnesium Requirement for 0.8x Scrubber Feed Saturation, as a Function of Oxidation. Values Calculated by Equations 2 - 9.

Table 1, EFFECT OF SO_2 PRECIPITATION RATE ON $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ SATURATION
IN LIMESTONE/ H_2O SCRUBBER

Reproduced from
best available copy.

Inlet SO_2 = 3000 ppm, Flue gas O_2 = 4.6%, SO_2 absorption = 95%

	Apr. (a) 22-26	Sept. 3-6	Aug. 12-16	Sept. 23-27	Sept. 16-20
Hold tank residence time, min.	6	12	12	5.3	5.3
Hold tank vol. (slurry), liter	473	1090	1090	484	484
Scrubber gas velocity, m./sec.	2.8	2.4	2.4	2.4	2.4
SO_2 feed rate, Kg./hr.	2.6	2.3	2.3	2.3	2.3
Precipitation rate, mg. SO_2 /min.(liter)	64	26	24	64	64
Flyash	no	no	yes	yes	no
Total solids, %	7	10	8	15	10
MgO feed, g./Kg.limestone	67	62	62	73	73
<u>Average Feed Liquor Composition</u>					
SO_4 g./liter	19.2	38.8	40.1	35.2	36.4
SO_3	3.0	1.9	6.8	6.0	7.4
Ca	0.24	0.64	0.69	0.27	0.28
Mg	7.8	10.6	12.0	11.4	11.4
Cl	6.2	2.05	3.2	2.9	3.3
pH	5.5	6.2	6.0	5.6	6.0
Total oxidation, %	26	20	28	12	14
Saturation ^(b)	0.43x	1.27x	1.16x	0.45x	0.43x
Saturation ^(c)	0.4x.	1.3x	1.2x	0.5x	0.4x

(a) open hold tank for Apr. 22-26 run, all other runs made with sealed tank.

(b) calculated at 50 C. with Reichtel-modified Radian equilibrium computer program.

(c) by solid $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ solubility test at 25°C.

as reduction of residence time to 5.3 min. resulted in saturation levels of 0.4x. This level of subsaturation was maintained with and without flyash addition. The choice of 5.3 min. residence time was made to provide the same precipitation rate that existed during the period of testing represented by the data of Figures 1 and 3. The precipitation rate is defined as:

$$\text{Precipitation rate} = \frac{\text{Amount of SO}_2 \text{ precipitated in hold tank}}{\text{Unit time (unit volume of slurry in hold tank)}}$$

where the amount of SO₂ precipitated in the hold tank is determined by the scrubber gas throughput, SO₂ inlet concentration, SO₂ scrubbing efficiency and oxidation. The value of the precipitation rate for the RTP scrubbers during unsaturated operation was 0.064 grams SO₂/min.(liter).

LIME VS. LIMESTONE

The effect of precipitation rate is probably the most important factor influencing the difference in CaSO₄·2H₂O saturation levels between lime and limestone scrubbers. The SO₂ precipitation rate is ultimately limited in the latter case by the comparatively slow dissolution rate of limestone. Lime scrubbers designed to take advantage of its rapid dissolution rate can operate in the unsaturated mode with hold-tank residence times at least as short as 1 minute. The high precipitation rates that ensue from these short residence times can explain the following advantages of lime over limestone scrubbers that have emerged from long-term concurrent TCA tests at the RTP facility:

- 1) Operating at 5000 ppm Cl⁻ and without MgO addition, the limestone scrubber averaged 1.2 - 1.3x saturation at 10 min. residence time, while the lime scrubber averaged 1.0x saturation at 2 min. residence time.
- 2) At 5000 ppm Cl⁻, the amount of MgO addition required to achieve a given level of subsaturation in the lime scrubber was about 1/10 that required for the limestone scrubber. The Mg⁺⁺ and dissolved sulfite levels are correspondingly lower in the lime system - an important consideration if subsaturated liquor is to be used for washing mist eliminators where oxidation will result in a rise in the local saturation level.

Another important advantage is the lower oxidation that is characteristic of the lime system, which is only 4% in our TCA scrubber at L/G 50 in the absence of Cl⁻. The lowest oxidation we have achieved in the limestone scrubber is about 10%. With Cl⁻ present these minima are about 12% in both cases. As would be expected, the chloride level at which a lime scrubber can operate without becoming supersaturated is much higher than that for a limestone scrubber, when no MgO is fed. In summary, the unsaturated mode is easier to achieve and maintain in a lime scrubber than in a limestone scrubber for any given set of operating conditions.

FACTORS AFFECTING OXIDATION

Low pH increases oxidation. Our experience supports this conclusion for both lime and limestone systems. The pH is, in turn, affected by chloride. In systems operating without chloride, where the scrubber effluent pH's are similar (5.6-5.7), the lime scrubber yields lower oxidation than the limestone scrubber. This follows from the higher pH and correspondingly lower concentration of dissolved sulfite throughout the lime system.

Addition of chloride lowered pH throughout the limestone scrubber. In the lime scrubber, where feed pH is controlled at a set value, chloride caused a greater drop in pH to occur across the tower resulting in a scrubber effluent pH lower (4.8-5.0) than the limestone scrubber (5.4-5.5). This drop in pH no doubt results from the common ion effect of Ca^{++} on the solubility of CaSO_3 , the dissolution of which is responsible for the SO_2 scrubbing action in lime systems operating in the tail-end-addition mode. As a result, lime scrubbers operating with chloride must use higher L/G's to maintain a given scrubbing efficiency and exhibit higher oxidation than lime scrubbers operating without chloride. For example, a scrubber with 4% oxidation at L/G 50 without chloride showed 12% oxidation at L/G 77 with 5000 ppm Cl.

Oxidation in the hold tank has been shown to contribute significantly to overall oxidation in the RTP scrubbers. Consequently, the hold tanks were sealed for all experiments related to unsaturated operation. When no MgO is fed to the scrubber, oxidation occurring in the filter/clarifier loop is of no consequence, because the rate of SO_2 circulation through that loop in the liquid phase is insignificant compared to the total SO_2 absorption (and purge) rate. Thus, even if totally oxidized, the SO_2 circulating through the filter would not affect the overall oxidation values. When MgO is added to either a lime or limestone system, the equilibrium concentration of dissolved sulfite is increased (primarily as MgSO_3°) and oxidation in the filter/clarifier can no longer be ignored. For example, operating at 10 g. Mg /liter and 5% solids, the total dissolved SO_2 was 1.8 g./liter in the hold tank. After passing through the filter loop this SO_2 was over 90% oxidized before reentering the hold tank. This amounted to 11% of the total SO_2 absorption rate in the scrubber and 30% of the total oxidation in the system. Raising the percent solids proportionately lowers the rate of circulation of slurry through the filter/clarifier loop and operation at 15% solids is recommended to limit this source of oxidation. This can be further reduced by filtering only settled sludge.

Oxygen in the flue gas averaged about 4.5% in this work, representing a coal-fired power plant at normal excess air. Increasing the oxygen content of the flue gas raised the oxidation rate in the scrubber tower and, although the limestone scrubber has been operated unsaturated with up to 6% oxygen, it appears that unacceptable levels of oxidation may be encountered at higher levels, especially in limestone/ MgO systems.

Flyash had no demonstrable effect on oxidation in either the lime or limestone scrubber.

L.G. & E. SIMULATION

In accordance with the rationale outlined above, the principal features of the Louisville Gas & Electric carbide-lime scrubber were simulated at RTP with the 3-stage TCA at 11.5 ft./sec. gas velocity, 3000 ppm inlet SO_2 and no Cl^- . Using ordinary pebble lime feed, which was slaked in filtrate liquor, the system ran unsaturated at 0.3x with only 100-200 ppm Mg in the scrubbing liquor (no MgO was added). As long as chloride was not present in the scrubber, this saturation level could be maintained indefinitely. It was concluded that carbide lime was not necessary to achieve the low oxidation characteristic of this mode, which in our case was only 4% at L/G 50. A sealed hold tank with 2-min. residence time was used in which the pH was controlled at 7.5. Additional tests with this tank in series with a larger, 20-min. open hold tank showed that complete precipitation of the sulfite (to 60 ppm) was achieved in the smaller tank; the large tank served no purpose as far as liquor saturation was concerned.

Addition of chloride to this system sharply dropped the scrubber effluent pH and the SO_2 scrubbing efficiency could be maintained only by raising L/G to 77. The scrubber effluent pH was not influenced by increasing the feed pH at this make-per-pass (560 mg. SO_2 /liter). The oxidation increased and the system became saturated. Unsaturated operation at a 5000 ppm level of chloride was again achieved by adding MgO to the lime feed to provide 1000-2000 ppm Mg^{++} in the scrubbing liquor.

SUPPLEMENTARY EFFECTS OF MgO ADDITION

Improvement in SO_2 scrubbing efficiency as a result of the accumulation of MgSO_4^0 and MgSO_3^0 in the scrubbing liquor, and/or the dissolution of MgO in the tower, is a generally recognized fact that is confirmed by our experience with both lime and limestone scrubbers. For example, operating the limestone scrubber as a multigrad tower (4 grids, 65% open area) gave 95% SO_2 removal with only 3.5 in. water ΔP at L/G 85, and 90% removal at L/G 65. The option of eliminating another reliability problem - TCA sphere deterioration - is thus indicated. These levels of efficiency required MgO feed rates of 1.7 lb/100 lb limestone. TCA scrubbers operating at normal pressure drops can achieve 98% removal under these conditions.

The high equilibrium concentration of sulfite in liquor containing large amounts of MgSO_4^0 , makes possible both rapid liquid-phase oxidation and rapid dissolution of solid CaSO_3 . Using a pressure oxidizer designed by M.W. Kellogg Co., we were able to increase oxidation from 28% in the scrubber slurry to 99% in the product sludge stream. As a result of this oxidation, the settling rate was increased from 0.07 to 0.8 cm/min and the final settled volume reduced from 340 to 190 ml per liter of original slurry. The final settled density was 54% and the volume of dry solid product was reduced by 40%. It is our expectation that an unsaturated limestone/ MgO scrubber can be operated with such an oxidizer to achieve fully oxidized product without affecting subsaturation in the scrubber loop.

Magnesium must be added as MgO either as calcined dolomite or half-calcined dolomite (to limestone scrubbers). It has been shown that the MgCO_3 normally present in limestones is not soluble at the pH levels at which these scrubbers operate. This magnesium enters and leaves the scrubber as the mineral dolomite, $\text{MgCO}_3 \cdot \text{CaCO}_3$, without participating in the scrubbing reactions. In our case, feeding limestone containing 5% MgCO_3 yields about 500 ppm Mg in the scrubber liquor - about 10% of the level that would be obtained by complete dissolution of the dolomitic component. The maximum CaCO_3 utilization obtainable from limestones of high Mg content is necessarily reduced because an equal amount of CaCO_3 is tied up in this insoluble mineral form.

CONCLUSIONS

- Unsaturated operation is technically feasible in either lime or limestone scrubbers. This feasibility has apparently been demonstrated on full scale systems at Louisville G. & E. and Mitsui.
- For any given set of operating conditions, subsaturated liquor is easier to achieve in lime than in limestone scrubbers.
- The presence of chloride makes it more difficult to operate in the unsaturated mode.
- Addition of magnesium makes it easier to operate unsaturated. The magnesium concentration can be increased by adding MgO.
- High SO_2 scrubbing efficiencies can be achieved at low pressure drop when magnesium concentrations of 10 g./liter or more are maintained in the scrubbing liquor.
- Minimizing oxidation assists in maintaining unsaturated conditions.
- Short hold tank residence times (1-2 min.) can be used in lime scrubbers operating in the unsaturated mode.

ACKNOWLEDGEMENTS

The RTP pilot plant from which these data were obtained is operated for EPA by Monsanto Research Corp. The contributions of this excellent group, and of Mr. Robert Opferkuch in particular, are gratefully acknowledged. Helpful advice and comments on this paper were received from Dr. Michael Epstein of Bechtel Corp. and from Frank Princiotta and John Williams of EPA. The computer program used for liquor saturation calculations was kindly provided by Charles Leivo of Bechtel.

NOMENCLATURE

L/G	liquid to gas ratio in tower, gal./1000 cfm @ 125°F
x	times saturation
α	activity coefficient of Ca^{++} and $\text{SO}_4^{=}$ in solution
μ	apparent ionic strength, assuming complete ionization of dissolved CaSO_4 and MgSO_4
c	empirically derived factor for expressing activities in terms of total dissolved concentrations
ratio	moles sulfate per mole sulfite in scrubber solids
oxidation	total moles SO_2 oxidized per mole SO_2 absorbed
CaSO_4°	ion complex formed by Ca^{++} and $\text{SO}_4^{=}$ in solution
Ca	total dissolved calcium in scrubber liquor, g./liter
MgSO_4°	ion complex formed by Mg^{++} and $\text{SO}_4^{=}$ in solution
Mg	sum of the concentrations of Mg^{++} and MgSO_4° , MgSO_3° excluded. g./liter

**LIMESTONE AND LIME TEST RESULTS AT THE
EPA ALKALI SCRUBBING TEST FACILITY
AT THE TVA SHAWNEE POWER PLANT**

by

M. Epstein, L. Sybert, S. C. Wang, C. C. Leivo and R. G. Rhudy

BECHTEL CORPORATION
50 Beale Street
San Francisco, California 94119

ABSTRACT

The limestone and lime reliability testing at the Shawnee facility has shown that scrubber internals can be kept relatively free of scale if the sulfate (gypsum) saturation of the scrubber slurry is kept below about 135 percent. This can be accomplished with the proper selection of percent solids recirculated, effluent residence time and liquid to gas ratio. The only significant reliability problem at the Shawnee facility has been associated with the scaling and/or plugging of the mist eliminator systems. The TCA utilizes a Koch Flexitray with underside steam sparging, in series with a chevron mist eliminator which has provision for underside washing. In a test run which is currently in progress, this mist elimination system has remained essentially clean over a 790 hour period, at a superficial TCA gas velocity of 8.6 ft/sec and 14 percent solids recirculated. The mist eliminator underwash utilizes diluted clarified liquor (at a ratio of three parts raw water makeup to two parts clarified liquor) at 0.3 gpm/ft² and the Koch tray is irrigated with the remaining clarified liquor plus mist eliminator wash at 0.5 gpm/ft². The spray tower utilizes a chevron mist eliminator with provision for underside and topside washing. Operability of this mist elimination system during tests of up to three months has been demonstrated, at a superficial gas velocity of 6.7 ft/sec and 8 percent solids recirculated, with intermittent or continuous bottomsides washing. A recent test with intermittent topside and bottomsides washing has indicated that operability of this mist elimination system can be extended substantially beyond three months.

Presented at the
SYMPOSIUM ON FLUE GAS DESULFURIZATION
Atlanta, Georgia
November 4-7, 1974

TABLE OF CONTENTS

<u>Section</u>	<u>Page</u>
1 INTRODUCTION	243
2 LIMESTONE/LIME RELIABILITY TESTING CONCEPTS	252
2.1 Scrubber Internals Scaling Potential	252
2.2 Mist Eliminator Scaling Potential	255
2.3 Closed Liquor Loop Operation	256
2.4 Determination of Sulfate (Gypsum) Saturation	257
3 LIMESTONE RELIABILITY TEST RESULTS	259
3.1 Performance Data and Test Evaluation	259
3.2 Conclusions	278
4 LIME RELIABILITY TEST RESULTS	281
4.1 Performance Data and Test Evaluation	281
4.2 Conclusions	298
5 PARTICULATE REMOVAL TEST RESULTS	301
6 REFERENCES	305

Section 1

INTRODUCTION

In June 1968, the Environmental Protection Agency (EPA), through its Office of Research and Development (OR&D) and Control Systems Laboratory, initiated a program to test a prototype lime and limestone wet-scrubbing system for removing sulfur dioxide and particulates from flue gases. The system is integrated into the flue gas ductwork of a coal-fired boiler at the Tennessee Valley Authority (TVA) Shawnee Power Station, Paducah, Kentucky. Bechtel Corporation of San Francisco is the major contractor and test director, and TVA is the constructor and facility operator.

The major goals of the test program are: (1) to characterize as completely as possible the effect of important process variables on sulfur dioxide and particulate removal; (2) to develop mathematical models to allow economic scale-up of attractive operating configurations to full-size scrubber facilities; and, (3) to perform long-term reliability testing.

The test facility consists of three parallel scrubber systems: (1) a venturi followed by a spray tower; (2) a Turbulent Contact Absorber (TCA); and, (3) a Marble-Bed Absorber. Each system is capable of treating approximately 10 Mw equivalent (30,000 acfm @ 300°F) of flue gas containing 1800-4000 ppm sulfur dioxide and 2 to 5 grains/scf of particulates. The test facility has been described in detail in References 1, 2 and 3.

The following sequential test blocks were defined for the program:

- (1) Air/water testing
- (2) Sodium carbonate testing
- (3) Limestone wet-scrubbing testing
- (4) Lime wet-scrubbing testing

The limestone and lime wet-scrubbing test blocks have been divided into three general categories: (1) short-term (less than 1 day) factorial tests, (2) longer term (over 2 weeks) reliability verification tests, and (3) long-term (4 to 6 months) reliability tests. The object of the factorial tests is to determine the effect of independent variables (e.g., gas rate) on SO_2 removal for the scrubber systems. The primary objective of the reliability verification tests is to define regions for reliable (e.g., scale-free) operation of the scrubber systems. The object of the reliability tests is to determine the long-term operating reliability for the scrubber systems and to develop definitive process economics data and scale-up factors. The test program has been described in detail in References 1, 2 and 3.

This report presents the results of the limestone and lime reliability tests at the Shawnee facility from October 1973 through October 1974. Results from the air/water, sodium carbonate, limestone factorial and limestone reliability verification testing have been presented in References 1, 2 and 3.

The lime reliability tests have been conducted on the venturi/spray tower system and the limestone reliability tests on the TCA system.* Figures 1 and 2 (drawn roughly to scale) show the venturi/spray tower and TCA scrubber systems along with the mist eliminator systems selected for de-entraining slurry droplets in the gas stream. The spray tower utilizes a chevron mist eliminator with provision for underside and top-side washing. The TCA employs a Koch Flexitray with underside steam sparging, in series with a chevron mist eliminator which has provision for underside washing.** Sections of the chevron mist eliminators used in the two scrubber systems are depicted, to scale, in Figure 3. The system configurations used for the venturi/spray tower lime reliability tests and the TCA limestone reliability tests are shown in Figures 4 and 5, respectively.

In June 1974, the EPA initiated a two-year Advanced Test Program at the Shawnee facility. The advanced program will involve, primarily, limestone testing on the TCA system and lime testing on the venturi/spray tower system. The major goals of the advanced program are:

- To continue long-term testing with emphasis on the demonstration of reliable mist elimination operation.

* Testing on the Marble-Bed system has been discontinued since July 1973 due to problems with erosion of the bed spray nozzles and subsequent pluggage of the marble bed.

** Underside mist eliminator washing was provided in July 1974 (see Section 3.1.8).

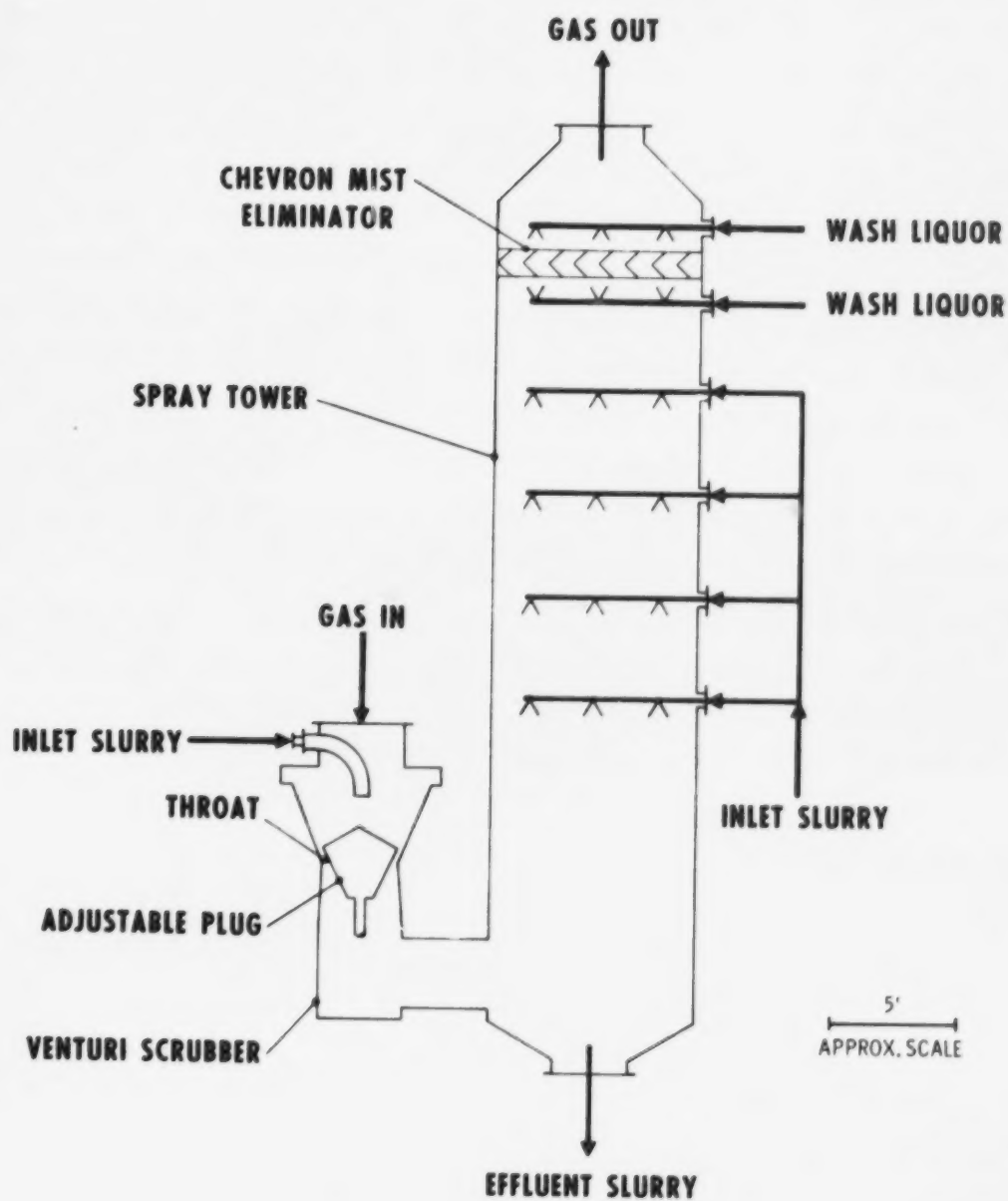


Figure 1. Schematic of Venturi Scrubber and Spray Tower

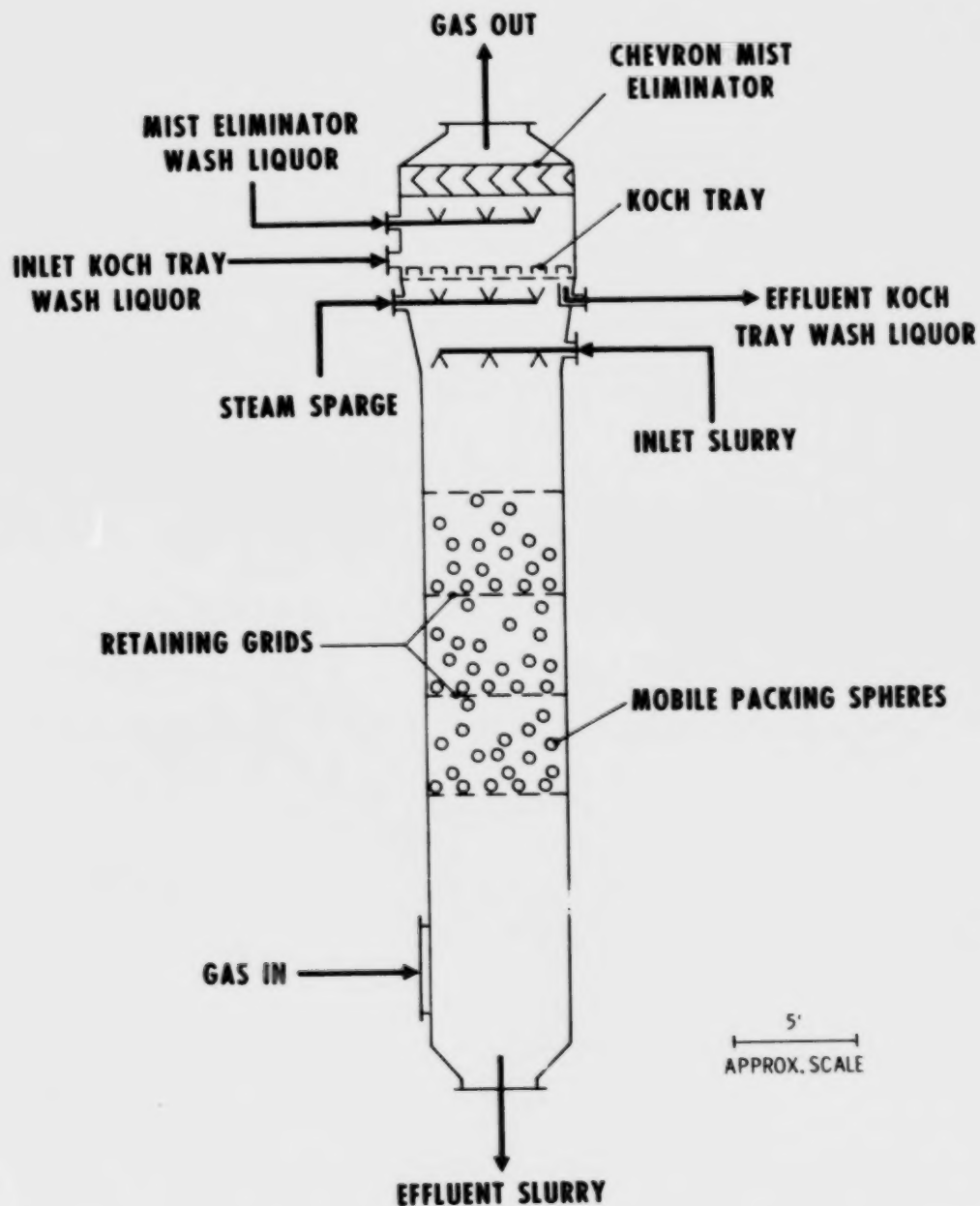


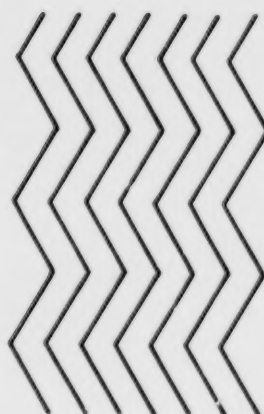
Figure 2. Schematic of Three-Bed TCA Scrubber

SPRAY TOWER



GAS FLOW

TCA



GAS FLOW

0 ————— 6 in.

Figure 3. Mist Eliminator Sections

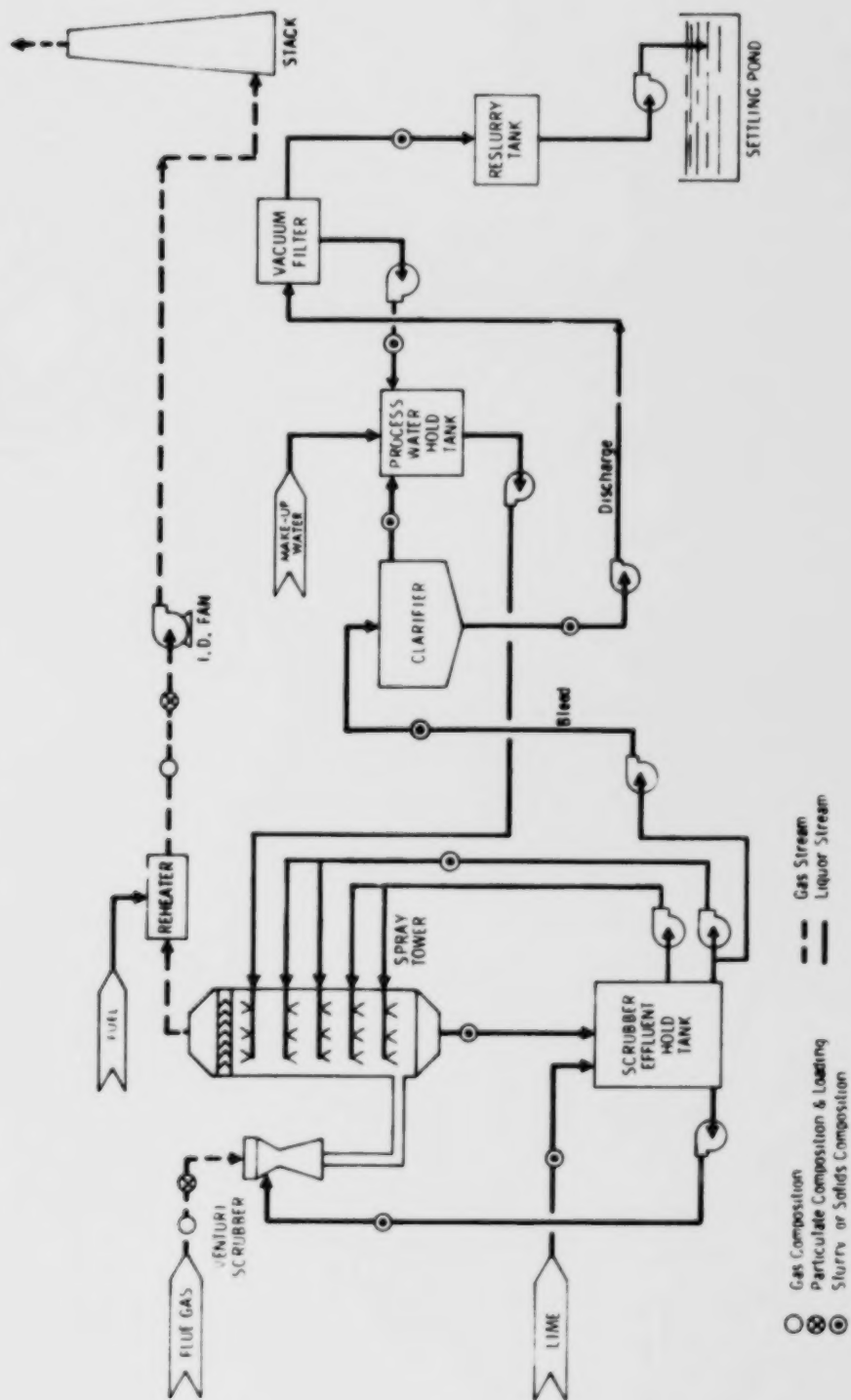


Figure 4. Typical Process Flow Diagram for Venturi Spray Tower System

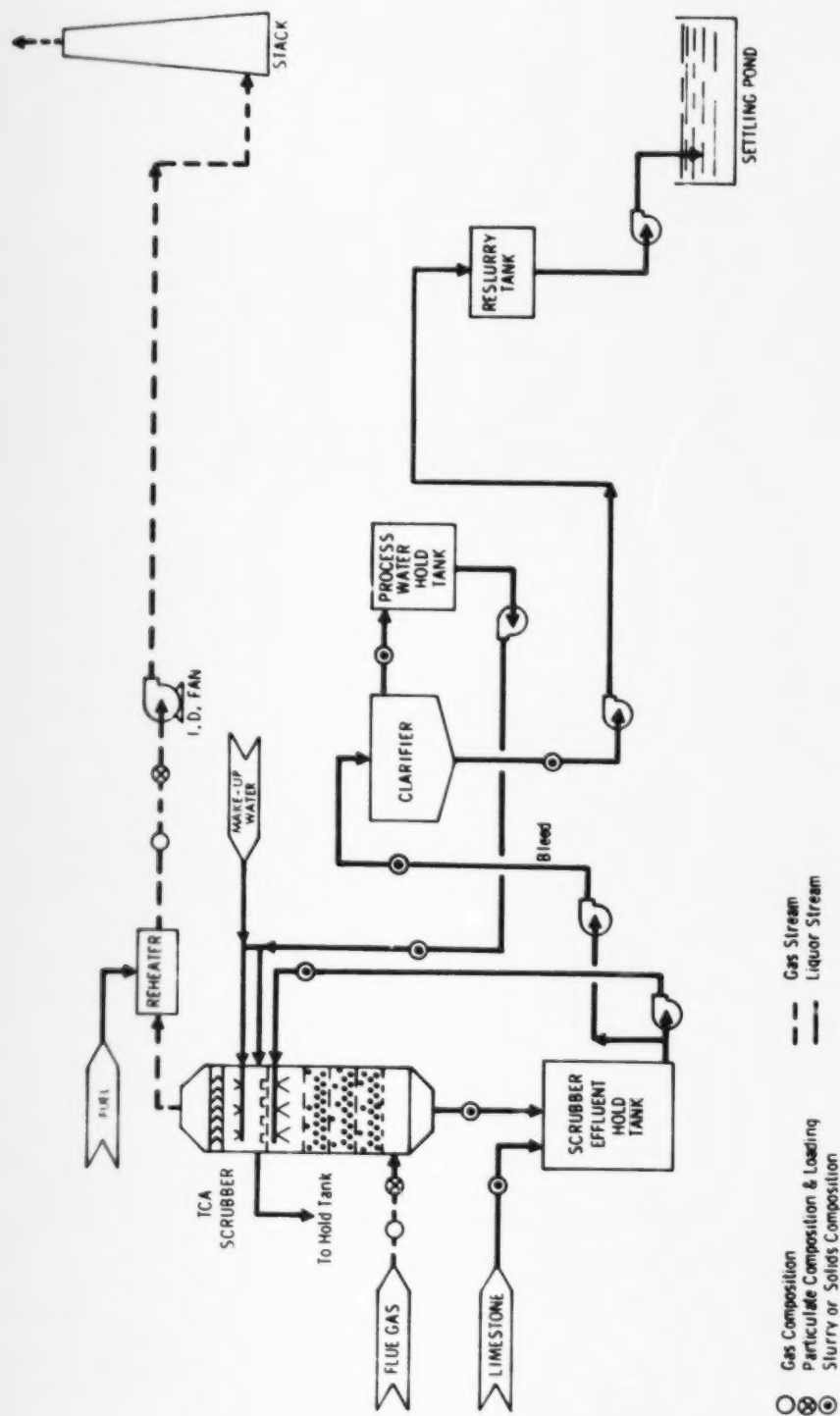


Figure 5. Typical Process Flow Diagram for TCA System

- To investigate advanced process and equipment design variations (e.g., operation with process slurry unsaturated with respect to gypsum) for the improvement of system reliability and process economics.
- To perform long-term (2 to 5 months) reliability testing on advanced process and equipment design variations.
- To evaluate process variations for substantial increase in limestone utilization with corresponding minimization of sludge production.
- To determine the practical upper limits of sulfur dioxide removal efficiency.
- To determine the effectiveness of existing technology for producing an improved throwaway sludge product.
- To evaluate system performance and reliability without fly ash in the flue gas.
- To evaluate the effectiveness of three commercially offered sludge fixation processes and untreated sludge disposal.
- To develop a computer program, in conjunction with TVA, for the design and cost comparison of full-scale limestone and lime systems.

The Advanced Test Program is described in detail in Reference 4. This manual, although not prepared for widespread distribution, is available upon request from Bechtel Corporation.

Section 2

LIMESTONE/LIME RELIABILITY TESTING CONCEPTS

As mentioned previously, the major objective of the limestone/lime reliability tests at the Shawnee facility is to determine long-term (4 to 6 months) operating reliability for the scrubber systems. Major reliability problems for limestone/lime wet-scrubbing systems have been associated with scaling and plugging* of scrubber internals and mist eliminator surfaces (a majority of mechanical problems can be solved by improved design).

2.1 SCRUBBER INTERNALS SCALING POTENTIAL

The Shawnee limestone reliability verification tests have shown that the scrubber internals can be kept relatively free of calcium sulfate (gypsum) scale at high percent solids recirculated and/or at high effluent residence times (see References 1 and 2). The results have also shown that scrubber scaling potential can be correlated with the calculated degree of sulfate saturation of the scrubber slurry, i. e., sulfate scaling is likely to occur for a degree of sulfate saturation

* In this report "scale" refers only to crystallizing hard solids, and "solids" or "soft-solids" refer to mud-like slurry solids. "Plugging" refers to the accumulation of mud-like slurry solids. Also, scale refers to sulfate base scale, unless otherwise noted.

of the scrubber inlet liquor greater than 130 to 140 percent.^{*} In Figure 6, the calculated sulfate saturations of the scrubber inlet liquor for the TCA limestone reliability verification tests are shown as a function of percent solids recirculated and effluent residence time.^{**} Figure 6 would indicate that, in order to remain below 130 percent saturated in a limestone system at 15 weight percent solids recirculated and at a liquid-to-gas ratio of from 64 to 80 gal/mcf, the effluent residence time would have to be equal to or greater than 10 minutes.

Results from the lime reliability tests at the Shawnee facility (to be discussed in Section 4) have shown that the calculated sulfate saturations of the scrubber inlet liquor can be similarly correlated with percent solids recirculated and effluent residence time. In addition, the lime testing has shown that the sulfate saturations are strong functions of flue gas inlet SO₂ concentration (SO₂ absorption rate).

Results from the EPA pilot facility at Research Triangle Park (Reference 6) have shown that it is possible to operate limestone and lime wet-scrubbing systems with liquors "unsaturated" with respect to calcium sulfate, thereby completely eliminating the potential for gypsum scaling. This is accomplished by controlling, within given limits, the total oxidation of sulfite to sulfate and the amount of dissolved sulfate.

^{*}The Radian Corporation (Reference 5) has determined that the "critical" sulfate saturation at 45°C is approximately 130 percent. The calculated sulfate saturations (at 50°C) in this report were made with the use of the Radian Equilibrium Computer Program (see Section 2.4).

^{**}TCA limestone reliability Run 526-2A (see Section 3) has been included in Figure 6.

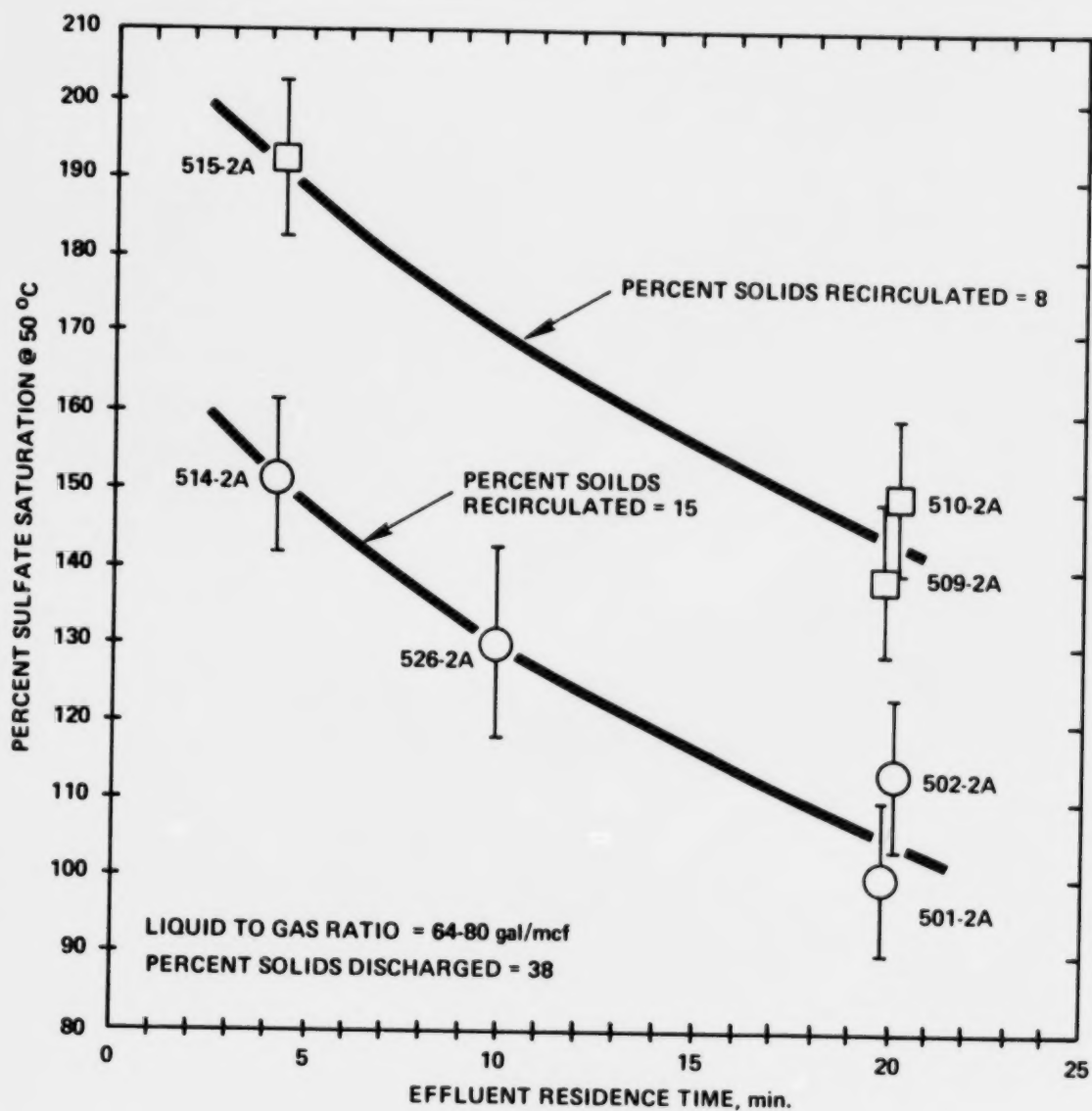


Figure 6. Calculated Degree of Scrubber Inlet Liquor Sulfate Saturation from TCA Reliability Verification Tests

Under these limits, the sulfate formed by oxidation is purged, without crystallization, as a "solid solution" within the precipitated CaSO_3 . The amount of sulfate which can be purged in this manner is related to the sulfate activity in the scrubbing liquor, which can be increased by addition of magnesium ion (see Reference 6). In order to keep the required magnesium addition as low as possible it is desirable to reduce overall oxidation. Oxidation may be reduced by (1) sealing the effluent hold tank (with a CO_2 or N_2 insulating blanket, if necessary), (2) adding oxidation inhibitors, and (3) reducing liquor rate. The required magnesium addition for unsaturated operation appears to be higher for limestone systems than for lime systems and also increases as the concentration of chlorides in the process liquor increases. The high concentration of chlorides in the Shawnee facility process liquors (see Tables 2 and 4 in Sections 3 and 4) are due to chlorides present in the coal which are converted to HCl and absorbed from the flue gas in the scrubber.

Operation in the unsaturated mode may also lead to improved process economics, as operating conditions conservatively selected for scale control under supersaturated liquor conditions can potentially be changed, e. g., effluent residence time and percent solids recirculated may be reduced. Moreover, the unsaturated process liquor may be used effectively for the irrigation of mist eliminators and wash trays.

2.2 MIST ELIMINATOR SCALING POTENTIAL

It is theorized that mist eliminator scaling is caused, predominantly, by SO_2 absorption into process liquor adhering to the mist eliminator surfaces. Calcium sulfate scale formation is subsequently caused by oxidation of the dissolved SO_2 and the resultant supersaturation of the liquor with respect to calcium sulfate. In addition, solids adhering to

the mist eliminator surfaces can cause pluggage and may also act as sites for additional scale formation. These problems may be alleviated, therefore, by: (1) reducing the sulfate saturation of the mist eliminator wash liquor (e.g., utilizing raw water or "subsaturated" process liquor); (2) washing the mist eliminator at a sufficient rate to remove all impacted solids; (3) utilizing a wash tray to reduce the suspended solids concentration of the droplets impinging upon the mist eliminator; (4) reducing the SO_2 concentration in the vicinity of the mist eliminator; and, (5) using mist eliminators with improved draining characteristics (e.g., sloped mist eliminators). All of these concepts will be tested at the Shawnee facility.

2.3 CLOSED LIQUOR LOOP OPERATION

Results at the Shawnee facility have shown that scaling potential (reliability) is significantly affected by the quantity of raw water input to the system, i.e., the greater the raw water input, the lower the scaling potential. In order to obtain significant reliability data, therefore, the scrubber systems must be operated in a "closed liquor loop" mode. Closed liquor loop operation is defined as operation wherein the raw water input to the system is equal to the water normally exiting the system in the settled sludge and in the humidified flue gas. Settled sludge densities in lime/limestone wet-scrubbing systems are normally equal to or greater than 38 percent by weight of solids. Closed liquor loop operation, of course, is also desirable from a water pollution standpoint.

The original test facility design included slurry pumps with water seals (Hydroseals) for bearing protection, water quench sprays for gas cooling

and dilute limestone slurry feed (10 - 20 wt% limestone). The water input under these conditions exceeded the makeup requirement for closed liquor loop operation. The systems operated, therefore, with partially open liquor loops during the limestone factorial tests, i.e., sludge with less than 40 wt% solids had to be discharged from the systems. This was not considered to be a serious problem during factorial testing for, at a specified scrubber inlet slurry pH, SO_2 removal is not significantly affected by liquor composition. However, little information was gained about the effect of scaling potential on reliability during this period.

The absorbent feed systems were changed in November 1972, to provide slurry feeds with up to 60 wt% limestone concentration. During the five week boiler outage in February and March 1973, the Hydroseal slurry pumps were converted to a Centriseal type (mechanical seal supplemented with air purge) and the TCA and Marble-Bed scrubbers were provided with process slurry gas cooling system. As a result of these modifications, closed liquor loop operation has been maintained at the facility since the beginning of limestone reliability verification testing in March 1973.

2.4 DETERMINATION OF SULFATE (GYPSUM) SATURATION

Three methods have been used at the Shawnee facility to determine the degree of sulfate (gypsum) saturation of the process liquors. These will be discussed below.

The most accurate of the three methods is a quantitative determination of the degree of sulfate saturation by the use of the Radian Equilibrium Computer Program (Reference 7) and the laboratory measured complete

liquor composition of a process liquor sample. The Radian Equilibrium Computer Program calculates the activities of calcium and sulfate ions, based on the measured liquor composition and pH. Hence, the degree of saturation is equal to the product of calcium and sulfate activities divided by the solubility product of gypsum at the specified temperature. The calculated sulfate saturations at 50°C in this report were based upon a solubility product for $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ of $2.2 \times 10^{-5} \text{ gmole}^2/\text{liter}^2$.

The second method is an approximate quantitative determination, which is less accurate but sufficiently rapid that it may be used in the field more easily to follow the scrubber performance. The test compares the concentrations of dissolved calcium and sulfate in the liquor sample with concentrations at saturation. The saturation concentrations are determined after the liquor sample is agitated with gypsum powder for a minimum of two hours. The degree of saturation is then calculated as the ratio of the product of calcium and sulfate concentrations before gypsum treatment to the product after treatment. This method assumes that the activity coefficients of calcium and sulfate remain unchanged after the gypsum treatment.

The third method is qualitative in nature, and does not involve any laboratory analyses or calculations. Stainless steel plates or strips are immersed in the process liquor in critical locations such as the effluent hold tank and Koch tray effluent surge tank, etc., to monitor the scaling potential of the scrubber system. These scale probes are examined periodically for scale formation. The scale growth rates can also be quantitatively determined by periodically weighing the probes. This method should be useful in a full scale facility, requiring a minimum of cost and operator attention.

Section 3

LIMESTONE RELIABILITY TEST RESULTS

Performance and analytical data from the TCA system limestone reliability testing at the Shawnee facility are presented in this section, along with an evaluation of each reliability test and the conclusions drawn, to date, from the testing.

3.1 PERFORMANCE DATA AND TEST EVALUATION

A summary of the test conditions and results for each of the TCA limestone reliability tests is presented in Table 1, along with the run philosophies. A summary of the scrubber inlet liquor analytical data for a majority of the tests is presented in Table 2, along with the calculated percent sulfate saturations. Essential operating data for all of the TCA limestone reliability tests are graphically presented in Reference 1. The operating data for the initial 480 hours of Run 535-2A, which is currently in progress, is graphically presented in Figure 7. An evaluation and discussion of each test is presented below.

3.1.1 TCA Run 525-2A

On October 24, 1973, the initial limestone long-term reliability test (Run 525-2A) was begun on the TCA system. The objective of the test

Table 1
SUMMARY OF LIMESTONE RELIABILITY TESTS ON TCA SYSTEM

Run No.	475-2A	525-2A	127-2A	528-2A
Start-of-Run Date	11/24/74	11/21/74	1/18/74	2/7/74
End-of-Run Date	11/15/74	1/10/74	1/24/74	2/25/74
On Stream Hours	617	1140	100	425
Gas Rate, acfm @ 100°F	25,000	25,000	20,000	20,000
Gas Velocity, fpm @ 120°F	15.5	8.5	8.5	8.5
Liquor Rate, gpm	1200	1200	1200	1200
L/G, gal/mcf	5.0	7.1	5.1	7.1
Percent Solids Recirculated	14.10	14.15.5	15.15	14.10
Effluent Residence Time, min	10	10	10	12
Stoichiometric Ratio, moles Ca added/mole SO ₂ absorbed	1.2-1.5	1.2-1.5	1.5-1.8	1.1-1.45
Avg % Limestone Utilization, 10% moles SO ₂ absorbed/mole Ca added	71	69	68	76
Inlet SO ₂ Concentration, ppm	1800-4000	1800-4000	2000-3600	2000-4000
Percent SO ₂ Removal	75-88	75-88	75-88	75-88
Scrubber Inlet pH Range	5.5-6.1	5.5-6.5	5.5-6.5	5.5-6.5
Scrubber Outlet pH Range	5.2-5.8	5.1-5.8	5.1-5.5	5.2-5.5
Percent Solids Discharged	15-30	15-30	15-30	15-30
Solids Disposal System	Clarifier	Clarifier	Clarifier	Clarifier
Loop Closure, % Solids Disch'd	51.42	50.47	54.59	55.15
Calculated % Solids Saturation in Scrubber Inlet Liquor @ 60°F	140	110	110	110
Total Dissolved Solids, ppm	2000-4000	2000-4000	2000-4000	4000-6000
Total AP Range Excluding Demister and Koch Tray, in H ₂ O	3.5-4.5	4.0-4.5	4.0-4.5	4.0-4.5
Demister and Koch Tray AP Range, in H ₂ O	2.0-2.7 (10-24 11/17/74) 2.5-3.1 (11/17-11/18)	1.9-2.1	2.1-3.1	2.3-2.8
Demister & P Range, in H ₂ O	5.0	5.15-5.25	5.0-5.15	5.15-5.25
Absorber	Limestone slurred to 10 wt % with makeup water and added to EHT.	Limestone slurred to 10 wt % with makeup water and added to EHT.	Limestone slurred to 10 wt % with makeup water and added to EHT.	Limestone slurred to 10 wt % with makeup water and added to EHT.
Mist Eliminator Koch Tray	No mist eliminator wash. K.T. 12 weir height irrigated with available makeup water (10 gpm) and all available clarified liquor (10 gpm). Bottom of Koch tray steam sparged 1 min/hr.	No mist eliminator wash. K.T. 12 weir height irrigated with available makeup water (10 gpm) and all available clarified liquor (10 gpm). Bottom of Koch tray steam sparged 1 min/hr.	No mist eliminator wash. K.T. 12 weir height irrigated with available makeup water (10 gpm) and all available clarified liquor (10 gpm). Not irrigated for 2 hr period. Bottom of Koch tray steam sparged 1 min/hr.	No mist eliminator wash. K.T. 12 weir height irrigated with available makeup water (10 gpm) and all available clarified liquor (10 gpm). Bottom of Koch tray steam sparged 1 min/hr.
Scrubber Internals	1 stage 4 grids with 4 spheres stage. Top bed-new TFB spheres. Middle 6 500 micron HDPE spheres.	1 stage 4 grids with 4 spheres stage. Top bed-new HDPE spheres. Middle 6 500 micron HDPE spheres from previous run.	2 stage 4 grids with 7 1/2 spheres stage. Top bed: 4 weir + 2 1/2 new HDPE spheres. Bottom bed-new HDPE's. Slurry nozzles not used (open pipe only) for this run only.	1 stage 4 grids with 4 spheres stage. All beds-new TFB spheres.
System Changes Before Start-of-Run	System for rubber. Koch tray and demister cleaned.	System cleaned. Replaced top bed spheres with new HDPE spheres. PAHT outlet Koch tray inlet pump switched from central to bottom bed.	System was cleaned. Removed top grid and bed of spheres. Middle bed: HDPE spheres from top bed + 2 1/2 new HDPE's. Bottom bed: 7 1/2 new HDPE's. Removed inlet slurry nozzles. Increased Koch tray weir height to 5.	System cleaned. Returned to 1 stage 4 grids with 4 new TFB spheres stage. Re-installed inlet slurry nozzles. Installed EHT's virginies in scrubber loop. Decreased Koch tray weir height to 2.
Method of Control	SO ₂ removal controlled at 84.2%	SO ₂ removal controlled at 84.2%	SO ₂ removal controlled at 84.2%	SO ₂ removal controlled at 84.2% Clarifier: Inlet pH 6.0-6.5 Stoich. Ratio 1.2-1.5
Run Philosophy	Intended long-term. Conditions chosen for long-term reliability based on reliability verification tests.	Intended long-term. Velocity reduced for reliability. HDPE spheres to be used until strainers are installed in scrubber loop.	Intended short-term. To determine whether changes in scrubber internal configuration could reduce slurry solids carryover.	Intended 2 wks. To observe any effect on demister solids and scale deposits with raw water only for Koch tray irrigation.
Results	Terminated due to rapidly increasing pressure drop across demister and Koch tray. Large solids deposits on underside of Koch tray and on scrubber walls between tray and main slurry header. TFB half-spheres plugged inlet slurry nozzles believed to be primary problem. Scrubber stages scale-free. Bottom of demister not plugged.	Terminated to replace worn HDPE spheres with new TFB spheres. Some solids accumulation between slurry nozzles and Koch tray due to partial blockage of spray nozzles with foreign materials. Bottom of demister not plugged.	Inconclusive. Solids carryover to Koch tray and demister appeared equal to or greater than previous run. Bottom of demister not plugged.	Demister and Koch tray areas clean. Due to less tight liquid loop from weepage of Koch tray raw water, calculated sulfate water tank was slightly lower at 110.

Table 1 (continued)
SUMMARY OF LIMESTONE RELIABILITY TESTS ON TCA SYSTEM

Run No.	529-2A	530-2A	531-2A	532-2A
Start-of-Run Date	2/26/74	3/28/74	4/10/74	7/17/74
End-of-Run Date	3/7/74	4/17/74	4/26/74	7/29/74
On Stream Hours	213	47	1088	288
Gas Rate, acfm @ 100°F	20,000	20,000	20,000	20,000
Gas Velocity, fpm @ 125°F	8.1	8.1	8.1	8.1
Liquor Rate, gpm	1200	1200	1200	1200
L/G, gal/mcf	74	74	74	104
Percent Solids Recirculated	14.5-15	14-15	15-9	7-8
Effluent Residence Time, min.	12	12	12	12
Stoichiometric Ratio, moles Ca added/mole SO_2 absorbed	2.35-1.95	1.2-1.55	1.5-1.7 (1.0-2.1), 1.19-2.0 (1.1-1.1), 1.0-2.5 (1.1-1.1)	1.1-1.5
Avg. % Limestone Utilization, 1000 moles SO_2 abso./mole Ca added	80	75	87 (100%), 81 (92%), 81 (124-190)	87
Inlet SO_2 Concentration, ppm	2300-3000	2200-4200	1700-1750	1900-3000
Percent SO_2 Removal	79-89	78-89	71-90	80-90
Scrubber Inlet pH Range	5.7-5.9	5.7-5.95	5.4-5.5	5.2-5.3
Scrubber Outlet pH Range	5.2-5.5	5.1-5.4	4.7-5.5	5.2-5.5
Percent Sulfur Oxidized	15-60	12-29	11-40	7-26
Solids Disposal System	Clarifier	Clarifier (centrifuge only after 4/10)	Clarifier	Clarifier
Loop Closure, % Solids Drying	90-97	90-93	90-92	92-93
Calculated % Sulfate Saturation in Scrubber Inlet Liquor @ 100°F	130	129	100	140
Total Dissolved Solids, ppm	4400	6000-6000	2,000	4000-6000
Total AP Range Excluding Demister and Koch Tray AP Range, in. H_2O	4.9-5.5	4.9-5.1	4.9-5.5	4.9-5.5
Demister and Koch Tray AP Range, in. H_2O	1.9-2.1	1.9-2.1	2.1-3.7	1.9-2.1
Demister AP Range, in. H_2O	0.19-0.23	0.17-0.28	0.14-0.16	0.15-0.19
Absorbent	Limestone slurried to 10 wt% with makeup water and added to EHT.	Limestone slurried to 10 wt% with clarified process liquor and added to EHT.	Limestone slurried to 10 wt% with clarified process liquor and added to EHT.	Limestone slurried to 10 wt% with clarified process liquor and added to EHT.
Mist Eliminator/Koch Tray	No mist eliminator wash. R.T. 12, weir height irrigated with available makeup water (1-2 gpm) and all available clarified liquor (1-2 gpm). Bottom of Koch tray steam-sparged 1 min/hr.	No mist eliminator wash. R.T. 12, weir height irrigated with 8 gpm makeup water and 7 gpm clarified liquor. Bottom of Koch tray steam-sparged 1 min/hr.	No mist eliminator wash. R.T. 15, weir height irrigated with available makeup water (1-2 gpm) and all available clarified liquor (1-2 gpm). Bottom of Koch tray steam-sparged 1 min/hr.	Recycle loop around mist eliminator R.T. 12. Bottom of mist elim. washed with 8 gpm makeup water plus 7 gpm recycle liquor. R.T. 12, weir ht. irrigated with 15 gpm recycle liquor plus 15 gpm mist elim. wash. Bottom of R.T. steam-sparged 1 min/hr.
Scrubber Internals	1 stage 4 grids with 4 spheres stage. All beds worn 1PB spheres from previous run.	1 stage 4 grids with 4 spheres stage. All beds worn 1PB spheres from previous run.	1 stage 4 grids with 4 spheres stage. All beds worn 1PB spheres from previous run.	1 stage 4 grids with 4 spheres stage. All beds worn 1PB spheres from previous run.
System Changes Before Start-of-Run	No cleaning. No changes.	System cleaned. Koch tray periphery sealed. PWRHT outlet (Koch tray) valve pump converted from beds to vented. Provided clean liquor for limestone slurry feed tank. Modified piping to obtain constant Koch tray makeup water liquor ratio. Sealed top of EHT.	System cleaned. Installed MgC add'n system. Lowered spray header 4'. Lowered Koch tray and further sealed periphery. Increased weir height to 1'. Installed inlet weir. EHT overflow line blanked.	System cleaned. Provided for closed-circuit recycle loop around mist eliminator Koch tray, including mist eliminator underwash. Lowered weir height to 2'. EHT and PWRHT sealed with CO_2 gas purge.
Method of Control	SO_2 removal controlled at 84.2%. Overrides: Inlet pH 5.5-6. Stoch. Ratio 1.0	SO_2 removal controlled at 84.2%. Overrides: Inlet pH 5.5-6. Stoch. Ratio 1.0	SO_2 removal controlled at: 81.0-82.1 84.2% 82.0-84.0 78.2% 84.0-85.0 84.2% 85.0-86.0 84.2%	Scrubber inlet pH controlled at 5.8-6.2
Run Philosophy	Intended long-term. Test conditions same as previous run, except Koch tray irrigated with all available makeup water and clarified liquor.	Intended long-term. EHT sealed to reduce oxidation, and Koch tray wash ratio held constant to observe any effect on solids buildup.	Intended long-term. Attempt to operate unsaturated with MgC add'n to sealed EHT. Percent solids lowered to 8% to provide additional clarifier liquor for Koch irrigation.	Intended long-term. Attempt to operate unsat. with MgC add'n to EHT. 10,000 ppm Mg. 1700 gpm liquor rate. EHT & PWRHT CO_2 blanket. mist elim. wash & mist elim. Koch tray recycle loop.
Results	Routine inspection after 213 hours revealed some soft solids and slight scale buildup on the demister lower vanes. Although amount of scale and solids was slight, it was considered to be significant after only 213 hours, and run was terminated. Bottom of demister 1PB plugged.	Unable to significantly reduce oxidation or sulfate saturation. Run terminated due to heavy scale and solids deposits on top of Koch tray, possibly caused by low (15 gpm) Koch tray flush rate and a clarifier rake malfunction on April 8 - 10, resulting in abnormal solids carry-over to Koch tray. Bottom of demister 445 plugged.	Unable to maintain unsaturated sulfate operation with 5000 ppm Mg. Steady state operation was not reached. Run terminated due to solids and scale buildup on mist eliminator, Koch tray and bottom grid. Bottom of demister 7PB plugged.	Unable to attain unsaturated sulfate operation with 10,000 ppm Mg. Run terminated due to scale build-up on Koch tray and loss of 4000 (w/ 2 beds) spheres through grids. Bottom of demister 1-2PB plugged.

Table 1 (continued)
SUMMARY OF LIMESTONE RELIABILITY TESTS ON TCA SYSTEM

Run No.	515-2A	514-2A	515-2A
Start-of-Run Date	8/6/74	9/1/74	9/12/74
End-of-Run Date	8/21/74	9/8/74	In Progress
On Stream Hours	142	100	-
Gas Rate, acfm @ 500°F	20,400	20,400	20,400
Gas Velocity, fpm @ 125°F	8.1	8.1	8.1
Liquor Rate, gpm	1200	1200	1200
L/G, gal/mcf	74	74	74
Percent Solids Recirculated	14 ± 1.5	10-12.5	12-15
Effluent Residence Time, min.	12	12	12/9-12-9-27, 15/after 9/27
Stoichiometric Ratio, moles Ca added/mole SO ₂ absorbed	1.1-1.55	1.10-1.40	1.35-1.70
Avg % Limestone Utilization, 100 moles SO ₂ abs./mole Ca added	74	82	65
Inlet SO ₂ Concentration, ppm	2900-3740	2700-3600	2000-4000
Percent SO ₂ Removal	95-99	75-80	78-88
Scrubber Inlet pH Range	5.7-6.0	5.4-5.85	5.7-6.0
Scrubber Outlet pH Range	5.5-5.85	5.2-5.4	5.1-5.5
Percent Sulfur Captured	15-21	15-10	15-10
Solids Disposal System	Clarifier	Clarifier	Clarifier
Loop Closure, % Solids Ditchg.	25-38	10-40	35-42
Calculated % Sulfur Saturation in Scrubber Inlet Liquor @ 500°F	135	110	120 (after 9/27)
Total Dissolved Solids, ppm	65,000-88,000	7500-8500	5000-8000
Total ΔP Range Excluding Demister and Knock Tray, in. H ₂ O	4.0-6.45	4.10-4.15	4.30-6.40
Demister and Knock Tray ΔP Range, in. H ₂ O	2.05-3.85	1.85-2.00	1.95-2.30
Demister ΔP Range, in. H ₂ O	0.15-0.25	0.15-0.20	0.10-0.20
Absorbent	Limestone slurried to 40 wt% with clarified process liquor and added to EHT.	Limestone slurried to 40 wt% with clarified process liquor and added to EHT.	Limestone slurried to 40 wt% with clarified process liquor and added to EHT.
Mist Eliminator Knock Tray	Recycle loop around mist eliminator, E.T. Bottom of mist elim. washed with 8 gpm makeup water plus 2 gpm recycle liquor. E.T. 12' water ht. irrigated with 15 gpm recycle liquor plus 15 gpm mist elim. wash. Bottom of E.T. steam sparged 1 min/hr.	Bottom of mist elim. washed cont. with 15 gpm 60L clar. liq. 100 gpm makeup water + 50 gpm clar. liq. E.T. irrig. with 40 gpm clar. liq. + 15 gpm mist elim. wash. Bottom of E.T. tray steam sparged 1 min/hr.	Bottom of mist elim. washed cont. with 15 gpm 60L clar. liq. 100 gpm makeup water + 50 gpm clar. liq. E.T. irrig. with 40 gpm clar. liq. + 15 gpm mist elim. wash. Total clar. liq. to mist elim. 6 E.T. 15 gpm min. Bottom of E.T. tray steam sparged 1 min/hr.
Scrubber Internals	4 stages of grids with 4 inches spheres/stage. All beds were TPR spheres from previous run.	4 stages of grids with 4 inches spheres/stage. All beds were TPR spheres from previous run.	4 stages of grids with 4 inches spheres/stage. All beds were TPR spheres from previous run.
System Changes Before Start-of-Run	System cleaned. Replaced worn TPR spheres with new TPR's.	System cleaned.	System cleaned. Seal water removed from cooling spray pump.
Method of Control	Scrubber inlet pH controlled at 5.8±0.2	Scrubber inlet pH controlled at 5.8±0.2	SO ₂ removal controlled at 85±5%. Overrides: Inlet pH 5.0, Stoch. Ratio 6:1, 1
Run Philosophy	Intended long-term. Attempts to operate unstart with 10,000 ppm Mg concn. in EHT. 1200 gpm liquor rate, 10% solids, CO ₂ blank over EHT & EHT, and mist elim. E.T. recycle loop.	Intended long-term. Discontinue use of MgO. Test conditions chosen similar to 515-2A.	Intended long-term. Test conditions chosen similar to 515-2A.
Results	Unable to attain unsaturated sulfate operation with 10,000 ppm Mg. Run terminated due to scale build-up on Knock tray and demister. Bottom of demister 12' plugged.	Run terminated due to scale build-up on the 12' tray. Bottom of demister 12' plugged.	Inspection of system at 78 hrs showed demister and Knock tray to be essentially clean. Heavy scale build-up on walls between steam spargers and slurry spray header. 10 mole scale in all grids.

Table 2

AVERAGE SCRUBBER INLET LIQUOR COMPOSITIONS AND CALCULATED SULFATE SATURATIONS FOR TCA LIMESTONE RELIABILITY RUNS

Run No.	Percent Solids Recirculated	Effluent Residence Time, min.	Percent Solids Discharged	Percent Sulfur Oxidized	Scrubber Inlet pH Range	Inlet Liquor Species Concentrations, mg/l (ppm)							Calculated Percent Sulfate Saturation at 50°C (d)	
						Ca ⁺⁺	Mg ⁺⁺	Na ⁺	K ⁺	SO ₄ ⁻²	SO ₃ ⁻²	CCl ₃ ⁻		
528-2A	14-16	10	31-42	15-30	5.5-6.1	2100	250	80	50	80	2000	70	3200	7000
529-2A	14-15.5	10	35-47	15-35	5.5-5.9	2300	140	50	140	80	1900	210	3700	8700
528-2A	14-16	12	25-33	19-30	5.7-5.95	1400	200	30	120	150	1600	250	1600	5400
529-2A (a)	14-15.5	12	10-47	15-30	5.2-5.5	1900	330	120	520	30	2000	180	3300	8400
530-2A (b)	14-16	12	10-43	12-25	5.1-5.4	1700	140	50	120	120	1900	320	2700	7370
532-2A	7-9	12	12-43	7-25	5.2-5.8	570	10,400	50	110	6000	18,700	30	2400	55,900
533-2A	14-15.5	12	29-38	15-23	5.7-6.0	715	11,400	50	110	2570	37,200	200	3100	55,750
534-2A	10-12.5	12	10-40	15-30	5.5-5.85	2220	170	40	180	70	1700	90	3540	8040
535-2A (c)	11-15.5	10	15-42	15-25	5.8-6.0	1800	290	50	70	50	1400	80	3050	7250

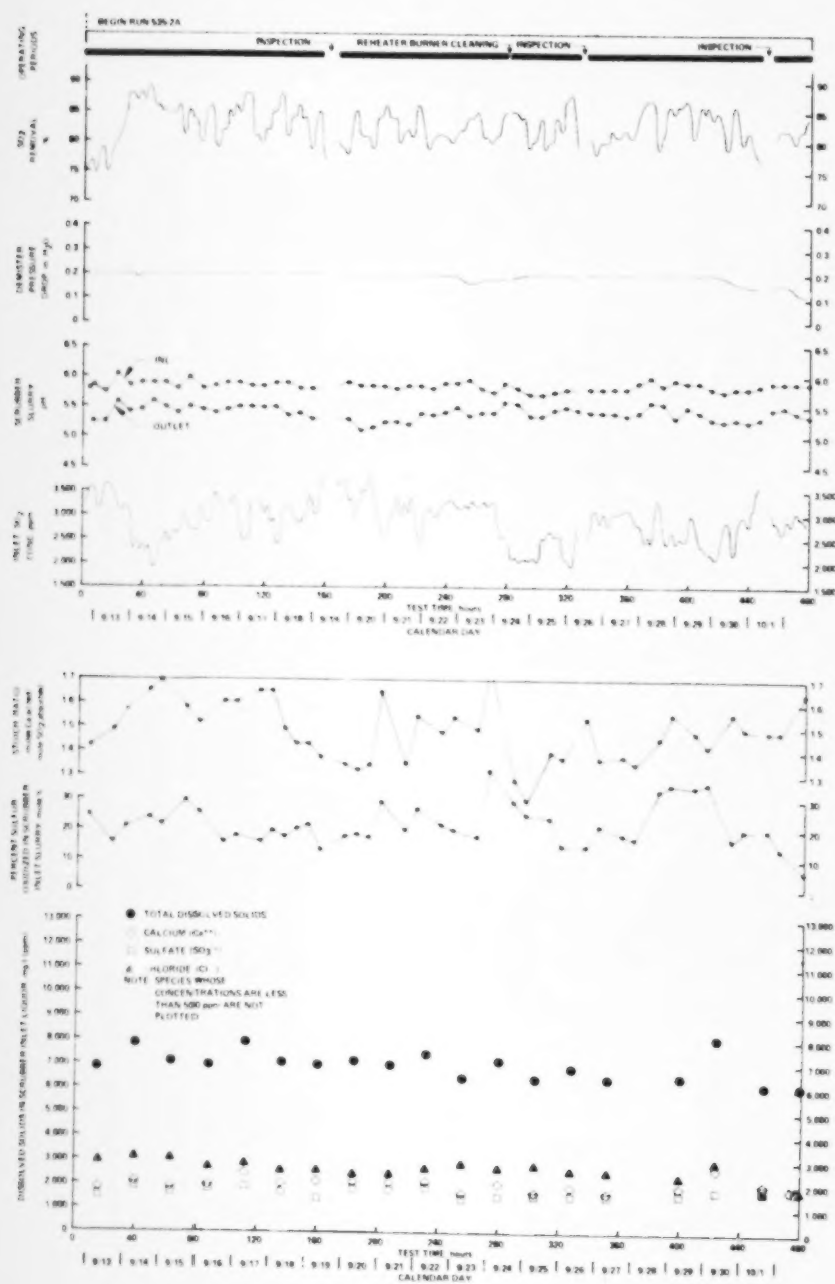
Note: Average concentrations are for steady-state operating periods. Solids disposal system: Clarifier only.

(a) Cells one sample was taken during steady-state operation.

(b) Solids disposal system: Clarifier only for first two-thirds of test, then centrifuge only.

(c) Test run was started on 9/12/74 with 12 minutes residence time. Residence time was increased to 15 minutes on 9/27. The run is still in progress as of 10/11. The values given in the table are for period 9/27 through 10/11.

(d) Calculated Sulfate Saturation = activity Ca⁺⁺ x activity SO₄⁻² / solubility product at 50°C. A solubility product for CaSO₄ · 2H₂O of 2.2×10^{-5} was used (Radco Corporation, A Theoretical Description of the Limestone-Injection Wet Scrubbing Process, NAPCA Report, June 9, 1979).



Gas Rate = 20,500 acfm @ 300 °F
 Liquid Rate = 1200 gpm
 L/G = 73 gal/mcf
 Gas Velocity = 8.6 ft/sec
 ENT (Soiled): Residence Time = 12 min (9/12-9/27),
 15 min (after 9/27)
 Three Stages, 5 in spheres/stage

Percent Solids Recirculated = 12.15 wt %
 Total Pressure Drop, Excluding Demister
 and Koch Tray = 4.0-4.6 in H₂O
 Scrubber Inlet Liquid Temperature = 120-126 °F
 Liquid Conductivity = 4,800-10,000 μ mhos/cm
 Discharge (Clarifier) Solids
 Concentration = 35.42 wt %

Figure 7. Operating Data for TCA Run 535-2A

was to operate continuously for from four to six months. The operating conditions selected for the run were based on the results of the reliability verification tests (see Figure 6) and the tests conducted at the EPA pilot facility at Research Triangle Park. The major test conditions were (see Table 1):

Gas Velocity	10.5 ft/sec
Liquid-to-Gas Ratio	60 gal/mcf
Percent Solids Recirculated	15
Effluent Residence Time	10 min.
Percent SC ₂ Removal (controlled)	84

During this test, the Koch Flexitray was irrigated with the available raw water makeup plus all the clarified liquor at an approximate one to three ratio, with the effluent irrigation liquor routed to the scrubber effluent hold tank. The mist eliminator, at that time, had no provision for underside wash and was washed with liquid entrainment from the Koch tray.

For the first time, thermo-plastic-rubber (TPR) spheres, supplied by UCP, were tested in the scrubber. They were used in the top bed while the bottom two beds were charged with high density polyethylene (HDPE) spheres, also supplied by UCP. The original wire mesh grids (which eroded during the limestone reliability verifications runs) had been replaced by sturdier bar-grids for the long-term reliability runs.

* The velocities listed in this report are based upon scrubber outlet gas flow rate at 125°F. Gas velocities listed in previous Shawnee reports have been based upon scrubber inlet gas flow rate corrected to 125°F. The velocities at the scrubber outlet conditions are approximately seven percent higher than at the scrubber inlet conditions, due to moisture pickup.

After 517 hours of operation the run was terminated, due to unusually heavy solids buildup on the underside of the Koch Flexitray and scale and solids buildup on the bottom vanes of the mist eliminator (about 60 percent of the free area was plugged). Numerous (over 200) half-spheres of the TPR type were found in the scrubber and slurry circulating system. TPR half-spheres were also found lodged in two of the four inlet slurry spray nozzles. It should be noted that the scrubber stages (and bottom-most grid) were essentially free of scale after the 517 hour operating period, as was expected, since the calculated scrubber inlet liquor sulfate saturation was 140 percent (see Table 2).

It is hypothesized that the accumulation of soft solids below the Koch tray was due, primarily, to the partial blockage of the slurry inlet nozzles, resulting in excessive entrainment of the fine slurry droplets. This excessive entrainment and the high gas velocity (10.5 ft/sec) most probably contributed to the mist eliminator pluggage.

3.1.2 TCA Run 526-2A

Run 526-2A was begun on November 21, 1973. The TPR spheres in the top bed had been replaced with HDPE spheres^{*}, and the accumulated scale and soft solids from Run 525-2A had been removed. The run conditions were identical to those for Run 525-2A, except that the gas velocity was reduced to 8.6 ft/sec. The velocity was reduced because more detailed investigation of previous reliability verification runs

* It was planned to use TPR spheres in all three stages following the installation of strainers in the inlet slurry piping.

(see References 1 and 2) indicated that long-term reliability for the present Koch tray/mist eliminator configuration could not be expected at a gas velocity of 10.5 ft/sec.

On January 10, 1974, the run was interrupted after 1190 hours of on-stream operation, to check the wear of the HDPE spheres in the three beds (HDPE sphere life had been estimated to be less than 2000 hours). Pressure drop across the chevron mist eliminator increased slightly during the initial 800 hours of operation, and during the last 400 hours increased more rapidly to a final level about 1.5 times the initial value of 0.18 inches H₂O.

The general appearance of the system was good. Scattered solids deposits (up to 1 inch) and light scale (about 1/16 inch) was found on the scrubber walls below the bottommost grid and on the wall areas not in contact with the spheres. The four bar grids were covered with 10-14 mil scale on the inactive surfaces (i.e., surfaces not in contact with spheres).

A heavy, relatively uniform solids layer covered the underside of the wash tray. All four inlet slurry spray nozzles were partially plugged with debris, primarily with plastic covering from pipe insulation. The flange of the steam sparger underneath the wash tray was found leaking. Mist eliminator plugging was confined to the bottom two passes only, reducing the free flow area by about 15 percent. Several small pieces of solids fell from the outlet gas duct and rested on top of the center section of the mist eliminator. The area restricted by these pieces was insufficient to affect the pressure drop across the mist eliminator.

As with Run 525-2A, it is hypothesized that the solids accumulations on the slurry inlet spray nozzles and header, on the adjacent scrubber walls and on the underside of the Koch wash tray were primarily caused by partial blockage of the TCA slurry inlet spray nozzles by debris, which changed the spray pattern and caused excessive entrainment of fine slurry droplets. Also, the blocked nozzles may have been a major factor contributing to the mist eliminator pluggage observed. It should be noted that the accumulation of sulfate scale on the bottommost bar grid (10-14 mils) was not excessive, after the approximate 1200 hours of operation. This was to be expected, since the scrubber inlet slurry was only 130 percent saturated with respect to calcium sulfate (see Table 2).

The average percent weight loss for the HDPE spheres in the bottom two beds was approximately 32 percent, for 1707 hours (Runs 525-2A and 526-2A) of use.

3.1.3 TCA Run 527-2A

Run 527-2A commenced on January 18, 1974, following the removal of the soft solids from the walls (inlet slurry spray nozzles area) and from the bottom of the Koch tray (the mist eliminator was not cleaned). It was a scheduled short-term run to determine whether certain changes in the scrubber internal configurations could minimize the mist eliminator pluggage problem and to observe the effect of the changes on SO₂ removal and limestone utilization. The modifications included raising the Koch tray outlet weir height from 2 to 3 inches, using two beds (3 grids) with 7 1/2 inches of spheres per bed and the removal of

the four inlet slurry spray nozzles for minimum spray generation with open pipe slurry distribution.

The results of the changes were inconclusive and the run was terminated after 133 operating hours. The slurry carryover to the Koch tray and the mist eliminator appeared to be equal to or greater than the carry-over observed during Run 526-2A.

The mist eliminator plugging continued on the lower vanes and the free flow area was reduced by an additional 50 percent (for a combined total of 65 percent for Runs 526-2A and 527-2A). It should be noted that the Koch tray, inadvertently, was not irrigated for a 2 hour period during the run.

An additional 20 mils of scale formed on the scrubber walls below the bottommost grid.

3.1.4 TCA Run 528-2A

Run 528-2A was started on February 6, 1974, following a complete cleaning of the system. The run was intended to be of short duration and was made using raw water only for Koch tray irrigation. The Koch tray effluent was routed to the sewer. The purpose of the run was to observe the effect of irrigation with raw water, versus raw water plus clarified liquor, on the mist eliminator scale and solids deposition. Scrubber internal configurations were returned to Run 526-2A conditions (2 inch Koch tray outlet weir height, use of inlet slurry spray nozzles, three-bed operation). The HDPE spheres were

replaced with new, improved, TPR spheres at 5 inches of spheres in each bed. A dual strainer (by Elliot Co.) was installed in the scrubber feed slurry loop to prevent the inlet spray nozzles from plugging by debris.

The run was terminated on February 26 after 425 hours of operation and the appearance of the system was generally good. The top of the Koch Flexitray was covered with approximately 20 mils of dust type solids. The mist eliminator was free of scale and solids with only a slight film of dust covering the inlet vanes. The pressure drop across the mist eliminator and Koch Flexitray had not increased during the test. The bottom of the wash tray was relatively clean.

Evidence from this test showed that the potential for scale and solids deposits on the mist eliminator is decreased when the degree of sulfate saturation of the Koch Flexitray wash liquor is reduced.

It should be noted that excessive weepage from the Koch tray irrigation water caused the test to be made with a slightly open liquor loop (average discharged solids of 29 percent). This explains the relatively low calculated scrubber inlet liquor sulfate saturation of 110 percent (see Table 2).

The run was interrupted for about 40 hours (February 11-13) due to a No. 10 boiler maintenance outage.

3.1.5 TCA Run 529-2A

Run 529-2A was started on February 26, 1974. The test conditions were the same as for Run 528-2A, except that the Koch tray was irrigated with the available raw water makeup plus all clarified liquor at an approximately one to three ratio. The run was initially intended to be a long-term limestone test. However, an inspection on March 7 revealed that the lower vanes of the mist eliminator were approximately 19 percent plugged with scale and soft solids. This was considered to be significant after the relatively short period of time and the run was terminated after only 213 operating hours.

Nearly all the TPR spheres were dimpled. * Eight spheres had failed (filled with slurry) in the bottom bed.

3.1.6 TCA Run 530-2A

Run 530-2A was started on March 28, 1974, after a thorough cleaning. The Koch tray was irrigated at a constant rate of 15 gpm, consisting of about 8 gpm makeup water and 7 gpm clarified liquor, with the purpose of achieving a high dilution of Koch tray irrigation liquor and, hence, minimizing the mist eliminator and Koch tray scaling potential (see Run 528-2A).

* This partially collapsed condition is considered normal by the supplier due to pressure and temperature cycling between operating periods and shutdowns.

The run was originally intended to be a long-term reliability test. However, it was terminated after only 476 operating hours when a rapid increase in pressure drop across the Koch tray occurred. An inspection revealed heavy scale and solids deposits on top of the Koch tray. A clarifier rake malfunction (which was not discovered until April 10) caused heavy solids carryover in the clarifier overflow for about two days. The solids carryover and the lower (15 gpm) Koch tray flush rate probably resulted in the settling of solids on the Koch tray. Subsequent use of the centrifuge for slurry dewatering resulted in centrate containing about 0.5 wt. % of solids, which may have contributed to further settling of solids on the Koch tray.

The mist eliminator was heavily plugged. Inspection on April 9 (after 290 hours of operation) showed an estimated overall restriction of 8 percent which then increased rapidly in 186 hours to 44 percent at the end of run.

Approximately 100 TPR spheres (0.4 percent of the total inventory of the system) failed during the run.

3.1.7 TCA Run 531-2A

Run 531-2A was begun on May 10, 1974, after a thorough cleaning. Magnesium oxide was added to the effluent hold tank in an attempt to operate in the sulfate unsaturated mode. Also, the TCA slurry header was lowered four feet in an attempt to reduce the amount of slurry droplet entrainment to the Koch tray.

The run was terminated after 1088 operating hours due to increasing pressure drop across the mist eliminator, Koch tray and bottommost TCA bed. Inspection revealed considerable scale and solids deposits on the Koch tray, mist eliminator and bottom grid. The mist eliminator was 75 percent plugged at the inlet edge and within the first pass. Scale deposits were limited to the first two passes.

Heavier than usual solids buildup was observed on the walls between the slurry spray nozzles and the steam sparger. This could be due to insufficient wetting of the walls by entrained slurry, resulting from the lowering of the slurry spray header.

The run had never attained steady state (i. e., the desired magnesium ion concentration of 10,000 ppm was not reached) and the final magnesium concentration of 5,000 ppm in the process slurry was insufficient to attain unsaturated operation.

A total of 638 TPR spheres (2.4 percent of the starting inventory) had failed during the run. The wear of the intact spheres varied from no measurable loss in the top bed to a 6.5 percent weight loss in the bottom bed for over 2000 hours of operation.

3.1.8 TCA Run 532-2A

Run 532-2A was begun on July 17, 1974, after the system was cleaned. Again, magnesium oxide was added to the effluent hold tank in an attempt to operate in the sulfate unsaturated mode at a 10,000 ppm level of magnesium ion in the process slurry. The scrubber recirculation liquor rate was increased from 1200 to 1700 gpm.

Also, a closed recycle irrigation liquor loop was provided for the Koch tray/mist eliminator system which included provision for mist eliminator underwash (~ 8 gpm fresh water plus ~ 7 gpm recycle liquor). The Koch tray recycle irrigation rate was set at 35 gpm. The clarifier liquor, which had been used for Koch tray irrigation during the previous tests, was routed directly to the effluent hold tank.

Because of relatively high saturation (140 percent) of the inlet scrubber slurry and the liquor in the wash tray loop, an inspection was made on July 29, after 258 operating hours. About 200 mils thick scale covered the walls below the bottommost grid and the top of the Koch tray was covered with 12 mils of scale. The free flow area of the mist eliminator was 1-2 percent plugged with solids, within the first and second passes.

About 4000 TPR spheres (15 percent of the total inventory) were found floating in the downcomer. They had passed through the bottom grid. *

In view of the above problems, the run was terminated. Note that, the SO_2 removal efficiency for this test averaged about 98 percent, as compared to 84 percent for the previous test, due to the addition of magnesium (see Table 1).

3.1.9 TCA Run 533-2A

Run 533-2A was started on August 6, 1974, after a thorough cleaning of the system. The recirculating slurry rate was reduced from 1700 to 1200 gpm, ** the used TPR spheres in the three beds were replaced

* The 1 1/2 inch TPR spheres can pass through the bar-grids supplied by UOP when they become dimpled on both sides.

** The liquor rate was reduced because of increased erosion of steel piping during the previous test.

with new ones and the percent solids recirculated was increased from 8 to 15 per cent. The Koch tray/mist eliminator system was the same as in the previous test, with a closed recycle irrigation liquor loop for the Koch tray and mist eliminator underwash. The scrubber slurry magnesium ion concentration was maintained at 10,000 ppm. The scrubber inlet liquor, however, was still supersaturated with respect to sulfate (see Table 2).

The run was terminated after 332 operating hours when an inspection showed that the underside of the wash tray was covered with up to 60 mils of scale over 40 percent of the tray area. The top surface of the tray was coated with from 50 to 100 mils of scale. The mist eliminator was 12 percent plugged with scale/solids deposit, primarily in the second and third passes. Also, the walls between the steam sparger and the slurry spray nozzles had up to 2 inches of solids deposit.

3.1.10 TCA Run 534-2A

Run 534-2A was started on September 3, 1974, after the system had been cleaned. It was decided to abandon, for the time being, attempts to operate in the sulfate unsaturated mode with the addition of magnesium oxide, until further exploratory tests were conducted at the EPA pilot facility in Research Triangle Park. Therefore, conditions for Run 534-2A were chosen similar to those of Run 526-2A, which had been in operation for 1190 hours, with no apparent scaling problem on the top of the Koch tray. The underside of the mist eliminator was washed continuously with 15 gpm diluted clarified liquor (~8 gpm makeup water plus ~7 gpm clarified liquor) while the Koch tray was

irrigated with the remaining ~ 8 gpm of clarified liquor plus the 15 gpm mist eliminator wash. The main differences between Runs 534-2A and 526-2A were, therefore, the addition of direct mist eliminator underwash and the four feet lower elevation of the inlet slurry spray header. In order to reach the expected steady state chloride concentration quickly, CaCl_2 was added to the system before startup.*

Run 534-2A was terminated after 100 on-stream hours due to scale buildup on the Koch tray. The top of the tray, including the valve caps, was almost completely covered with from 10 to 15 mils of scale. About 2 percent of the mist eliminator flow area was restricted by solids. Also, about 30 percent of the Koch tray underside surface was covered with solids, reaching a maximum thickness of 1 1/2 inches.

As in the previous run, the walls between the steam sparger and the slurry spray nozzles had up to 2 inches of solids deposit.

3.1.11 TCA Run 535-2A

Run 535-2A was begun on September 12, 1974, after the system had been cleaned. The test conditions were nearly identical to Run 534-2A, with the added condition that a minimum clarified liquor flow rate of 15 gpm for Koch tray irrigation and mist eliminator wash was to be

* It normally takes a number of weeks for the chloride level in the process slurry to reach a steady state value, when beginning a run with raw water. The process slurry was dumped at the beginning of Run 534-2A to dispose of the large quantity of magnesium ion in the system.

maintained. In maintaining this rate, the fluctuating bleed flow to the clarifier resulted in, at times, only 12-13 percent solids (as compared to the desired 15 percent) in the recirculating slurry. Therefore, the effluent residence time was increased from 12 to 15 minutes, in order that the sulfate saturation of the scrubber liquor remain below 135 percent at all times (see Figure 6). Also, about 12 percent more make-up water was available for mist eliminator/Koch tray irrigation, because of the elimination of seal water from a flue gas cooling spray pump.

After 158 hours of operation, an inspection showed about 15 mils of scale covering about 25 percent of the top of the Koch tray and relatively heavy solids accumulation (up to two inches thick in some areas) on the underside of the tray. The mist eliminator was about 5 percent plugged with solids (mostly fly ash) on the lower vanes. The walls between the steam sparger and the slurry spray nozzles had up to 4 inches of solids deposit. The run was continued.

After 316 hours of operation, a second inspection revealed that the scale had disappeared from the top of the Koch tray and that the mist eliminator was only 1-2 percent plugged with scattered solids (mostly fly ash) on the lower vanes. The bottom of the Koch tray, however, still had heavy solids accumulation. At this time, it was found that the steam pressure in the Koch tray steam sparger (see Figure 2) had drifted to 50 psig from an original setting of 125 psig. The steam pressure was reset and the run continued.

After 431 hours of operation, a third inspection showed that the top of the Koch tray was still clean and that the mist eliminator was

practically clean (approximately one percent plugged with scattered solids on the bottom vane). The solids accumulation on the underside of the Koch tray had been substantially reduced, with 30 percent of the underside clean to metal. However, solids buildup on the walls between the steam sparger and the slurry spray nozzles had increased to a maximum thickness of 8 inches in some areas.

After 786 hours of operation, a fourth inspection showed that the top of the Koch tray was still clean and the mist eliminator again practically clean (approximately one percent plugged with scattered solids on the bottom vane). The solids accumulation on the underside of the Koch tray was slightly increased, with 20-25 percent of the underside clean. The solids buildup on the walls between the steam sparger and the slurry spray nozzles had further increased. The maximum deposit in one area had reached 15 inches in thickness. Approximately 30 mils of scale had accumulated on the grids during the 786 hours of operation. There was no measurable increase in pressure drop across the beds, Koch tray or mist eliminator. The run was continued following the 786 hour inspection.

The average wear of the TPR spheres, for 1200 hours of operation, was about 4 percent.

3.2 CONCLUSIONS

3.2.1 Scrubber Internals Operability

Both the limestone reliability verification and reliability test results have shown that scrubber internals can be kept relatively free of scale

if the sulfate (gypsum) saturation of the scrubber liquor is kept below about 135 percent. This can be accomplished with the proper selection of percent solids recirculated, effluent residence time and liquid-to-gas ratio (see Figure 6). The data indicate that, at approximately 15 percent solids recirculated, 10 minutes effluent residence time and 73 gal/mcf liquid-to-gas ratio, the scaling of the TCA internals will not become a problem to maintaining long-term system operation. For example, the four bar grids in the TCA were covered with only 10-14 mils of scale after 1190 hours of operation in Run 526-2A.

A significant recent problem in the TCA has been associated with solids buildup on the walls between the steam sparger and the slurry spray nozzles (see Runs 531-2A through 535-2A), which originated after the slurry spray nozzles had been lowered by four feet. It is hypothesized that the solids buildup is due to insufficient wetting of the walls by entrained slurry. The problem may be corrected, therefore, by washing the walls between the tray and nozzles or by raising the nozzles to their original position.

Until recently, a significant limiting factor in the long-term reliability of the TCA scrubber has been associated with the erosion and subsequent collapse of the UOP supplied HDPE spheres (sphere life was approximately 2000 hours). The collapsed spheres eventually filled with slurry and settled to the bottom of the support grids. Starting with Run 528-2A, an improved type of UOP supplied TPR sphere was used in the TCA beds. After approximately 2500 hours of operation, all the spheres were dimpled on one side, about 2.4 percent failed at the seam and the weight loss averaged about six percent. It is estimated that these TPR spheres would not need replacement until after about a year of service. The problem associated with dimpled spheres falling through the bar-grids (see Run 532-2A) can be solved by redesigning the grids.

There has been no evidence of significant erosion of the bar-grids in the TCA after more than 5000 hours of operation.

3.2.2 Koch Tray/Mist Eliminator Operability

The most significant reliability problem encountered during the TCA limestone reliability tests has been associated with scaling and/or plugging of the Koch Flexitray and bottom vanes of the chevron mist eliminator (see Figures 2 and 3).

Run 528-2A showed that the mist eliminator and top surface of the Koch tray can be kept relatively free of scale if the irrigation liquor is low in sulfate saturation (unsaturated).

Run 526-2A showed that the top of the Koch tray can be kept relatively free of scale if it is irrigated with the available raw water makeup and clarified liquor, provided the mixture is maintained below the critical sulfate saturation level. The run was terminated prematurely, after 1190 hours, due to excessive wear on the HDPE spheres and pluggage (with debris) of the scrubber slurry nozzles.

The present Run 535-2A has been in operation for over 780 hours without any significant scaling or plugging of the top of the Koch tray or mist eliminator surfaces. The test conditions for this run are similar to those for Runs 526-2A and 529-2A, with the exception of direct mist eliminator underwash. It is likely that long-term reliability of this mist elimination system would be realized at the run conditions tested.

Section 4

LIME RELIABILITY TEST RESULTS

Performance and analytical data from the venturi/spray tower system lime reliability testing at the Shawnee facility are presented in this section, along with an evaluation of each reliability test and the conclusions drawn, to date, from the testing.

4.1 PERFORMANCE DATA AND TEST EVALUATION

A summary of the test conditions and results for each of the venturi/spray tower lime reliability tests is presented in Table 3, along with the run philosophies. A summary of the scrubber inlet liquor analytical data for a majority of the tests is presented in Table 4, along with the calculated percent sulfate saturations. ^{*} Essential operating data for all of the venturi/spray tower lime reliability tests are graphically presented in Reference 1. As an example, operating data for Run 604-1A from June 6, 1974, to June 24, 1974, is graphically presented in Figure 8. An evaluation and discussion of each test is presented below.

^{*} The sulfate saturations and analytical data are for an average inlet gas SO₂ concentration of 2800 ppm. The effect of inlet gas SO₂ concentration on sulfate saturation is discussed in the following sections.

Table 3

SUMMARY OF LIME RELIABILITY TESTS ON VENTURI/SPRAY TOWER SYSTEM

Run No.	601-1A			602-1A	603-1A
Start-of-Run Date	10/9/74			5/15/74	4/2/74
End-of-Run Date	1/8/74			4/1/74	4/19/74
On Stream Hours	2155			593	595
Gas Rate, acfm @ 530°F	25,000			25,000	25,000
Spray Tower Gas Vel, fps @ 125°F	6.7			6.7	6.7
Venturi Spray Tower Liquor Rates, gpm	600/1200			600/1200	600/1200
Spray Tower L/G, gal/mcf	6.0			6.0	6.0
Percent Solids Recirculated	7.9			7.5-9.5	13.5-16
Effluent Residence Time, min	12			12	12
Solids Disposal System	Clarifier Only (10/9-11/7)	Clarif. & Intermittent Filter (11/7-12/15)	Clarifier & Filter (12/15-1/8)	Clarifier & Filter	Clarifier & Filter
Stoichiometric Ratio, moles Ca added/mole SO ₂ absorbed	1.01-1.28	1.02-1.18	1.04-1.19	1.02-1.18	1.00-1.13
Avg % Limestone Utilization, 100% moles SO ₂ abs./mole Ca added	87	91	90	91	94
Inlet SO ₂ Concentration, ppm	1600-1900	1600-4000	2100-4400	2100-3800	2100-4300
Percent SO ₂ Removal	78-91	75-95	75-95	87-97	85-99
Scrubber Inlet pH Range	7.4-8.8	7.4-8.4	7.7-8.4	7.6-8.3	7.6-8.2
Scrubber Outlet pH Range	4.7-5.5	4.7-5.5	4.8-5.5	4.9-5.4	4.8-5.3
Percent Sulfur Oxidized	10-30	10-30	10-30	5-28	12-22
Loop Closure, % Solids Dischg.	20-26	20-27	42-52	42-48	46-54
Calculated % Sulfate Saturation in Scrubber Inlet Liquor @ 500°C	100	120	180	165	135
Dissolved Solids, ppm	5700-7400	4000-7400	9800-12,500	9500	8200-10,900
Total AP Range, Excluding Demister, in. H ₂ O	11.0-11.5	11.0-12.0	11.7-12.5	11.0-12.0	11.0-12.0
Venturi AP, in. H ₂ O	9	9	9	9	9
Demister AP, in. H ₂ O	0.55-0.70	0.55-0.90	0.40-1.05	0.55-0.73	0.55-0.72
Absorbent	Lime slurried to 20 wt% with makeup water and added to EHT.			Lime slurried to 20 wt% with makeup water and added to EHT.	Lime slurried to 20 wt% with makeup water and added to EHT.
Mist Eliminator	10/9 - 12/15: Bottom washed with available makeup water (~14 gpm) plus available clarified liquor (~25 gpm). Continuous wash rate of 0.8 gpm/ft ² . 12/15 - 1/8: Bottom washed with available makeup water (~4 gpm) plus available clarified liquor (~4 gpm). Wash rate of 1 gpm/ft ² on cycle of 1-1/2 min on 1-1/2 min off. Also, top was washed once wk with fresh water for 5 min during last 4 wks.			Bottom washed with available makeup water (~5 gpm) plus available clarified liquor (~14 gpm). Wash rate of 1 gpm/ft ² on cycle of 4 min on 1 min off.	Bottom washed with available makeup water (~5 gpm) plus available clarified liquor (~21 gpm). Wash rate of 1 gpm/ft ² on cycle of 1-1/2 min on 1-1/2 min off.
Scrubber Internals	All nozzles on 4 headers in top 3 headers sprayed downward. Bottom header nozzles (2) sprayed upward.			All nozzles on 4 headers sprayed downward. 7 nozzles/header on top 3 headers. 5 nozzles on bottom header.	All nozzles on 4 headers sprayed downward. 7 nozzles/header on top 3 headers. 6 nozzles on bottom header.
System Changes Before Start of Run	System, scrubber and mist eliminator cleaned.			System cleaned chemically (Na ₂ CO ₃ /sugar limestone) wash soln. followed by mech. cleaning. EHT sealed. Installed external combustion reheater. Changed nozzles on bottom spray header to spray downward instead of upward. Capped middle nozzles on bottom header.	Mist eliminator cleaned.
Method of Control	Scrubber inlet pH controlled at 8.0 ± 0.2			Scrubber inlet pH controlled at 8.0 ± 0.2	Scrubber inlet pH controlled at 8.0 ± 0.2
Run Philosophy	Initially started as lime reliability verification test. Subsequently, due to apparent reliability of the run, decision was made that test continue as long-term reliability test.			Intended long-term. Sealed EHT in attempt to reduce sulfite oxidation and thereby degree of sulfate saturation.	Intended long-term. Recirculate 15% solids in attempt to reduce degree of sulfate saturation. EHT sealed.
Results	Routine inspection on 11/7/74 showed system was generally clean after 600 hours of operation with clarifier only for solids disposal. Run was terminated on 1/8/74 due to ID fan vibration and rapidly increasing pressure drop across demister. Sulfate based scale formed on most scrubber walls and in slurry piping. Top of mist eliminator 80% plugged with solids that fell from outlet duct-work. Mist eliminator top-vanes heavily scaled (~500 mils avg.).			Run terminated due to scale (~125 mils avg.) and solids deposits on mist eliminator top-vanes. Sulfite oxidation and sulfate saturation were not reduced. Steady state operation not achieved.	Degree of sulfate saturation was reduced, but run was terminated due to scale (~8 mils avg.) and solids buildup on the mist eliminator top-vanes.

Table 3 (continued)

SUMMARY OF LIME RELIABILITY TESTS
ON VENTURI/SPRAY TOWER SYSTEM

Run No.	004-1A	005-1A	006-1A	008-1A
Start-of-Run Date	4/26/74	7/31/74	8/7/74	8/21/74
End-of-Run Date	7/15/74	8/6/74	8/14/74	9/17/74
On Stream Hours	1828	141	170	610
Gas Rate, acfm @ 110°F	25,000	25,000	25,000	25,000
Spray Tower Gas Vel. fpm @ 125°F	6.7	6.7	6.7	6.7
Venturi/Spray Tower Liquor Rates, gpm	min. 1-1000/1200	min. 1-1000/1200	min. 1-1000/1200	1000/1200
Spray Tower L/G, gal/mcf	<0	<0	<0	<0
Percent Solids Recirculated	7.5-9.0	8.0-9.5	7.7-9.0	7.7-9.4
Effluent Residence Time, min.	17	17	17	12
Solids Disposal System	Clarifier & Filter	Clarifier & Filter	Clarifier	Clarifier & Filter
Stoichiometric Ratio, moles Ca added/mole SO ₂ absorbed	1.03-1.10	1.10-1.17	1.10-1.15	1.03-1.17
Avg % Limestone Utilization, 100x moles SO ₂ abs./mole Ca added	88	88	89	88
Inlet SO ₂ Concentration, ppm	2000-1800	2500-3500	2400-3200	2000-3500
Percent SO ₂ Removal	70-92	74-81	74-79	74-95
Scrubber Inlet pH Range	7.7-8.4	8.8-9.2	7.8-8.2	7.9-8.4
Scrubber Outlet pH Range	4.5-5.4	4.9-5.1	5.0-5.2	4.8-5.1
Percent Sulfur Oxidized	8-10	12-28	12-22	13-24
Loop Closure, % Solids Dischg.	50-60	48-52	18-25	48-58
Calculated % Sulfate Saturation in Scrubber Inlet Liquor @ 50°C	130	115	120	130
Dissolved Solids, ppm	11,000-13,700	5,000-7,400	5,000-7,000	7,500-9,500
Total ΔP Range, Excluding Demister, in. H ₂ O	3.5-5.8	5.2-5.9	5.5-5.7	11.5-12.0
Venturi ΔP, in. H ₂ O	Plug 100% open	Plug 100% open	Plug 100% open	9
Demister ΔP, in. H ₂ O	0.70-1.75	0.80-0.85	0.78-0.86	0.75-0.92
Absorbent	Lime slurried to 20 wt % with makeup water and added to scrubber downcomer.	Lime slurried to 20 wt % with makeup water and added to scrubber downcomer.	Lime slurried to 20 wt % with makeup water and added to scrubber downcomer.	Lime slurried to 20 wt % with makeup water and added to scrubber downcomer.
Mist Eliminator	Bottom washed with available makeup water (~5 gpm) plus available clarified liquor (~35 gpm). Wash rate of 1 gpm/ft ² on cycle of ~3 1/2 min on/1 1/2 min off.	Bottom washed with available makeup water only (~5 gpm). Wash rate of 0.4 gpm/ft ² on cycle of ~1 min on/4 min. off.	Bottom washed continuously with 15 gpm (0.3 gpm/ft ²) raw water only. (Rate is greater than available makeup water).	Bottom washed with available makeup water only (~5.5 gpm). Wash rate of 150 gpm (3 gpm/ft ²) for approx. 9 min. every 4 hours.
Scrubber Internals	All nozzles on 4 headers sprayed downward. 7 nozzles/header on top 3 headers. 5 nozzles on bottom header.	All nozzles on 4 headers sprayed downward. 7 nozzles/header on top 3 headers. 5 nozzles on bottom header.	All nozzles on 4 headers sprayed downward. 7 nozzles/header on top 3 headers. 5 nozzles on bottom header.	All nozzles on 4 headers sprayed downward. 7 nozzles/header on top 3 headers. 5 nozzles on bottom header.
System Changes Before Start of Run	Mist eliminator and outlet duct cleaned. Sealed EHT provided with N ₂ gas purge. EHT overflow blanked. Lime slurry makeup added to scrubber downcomer.	System cleaned.	Mist eliminator cleaned.	Mist eliminator cleaned. Provided for Freon gas blanket over EHT.
Method of Control	Scrubber inlet pH controlled at 8.0 ± 0.2	Scrubber inlet pH controlled at 9.0 ± 0.2	Scrubber inlet pH controlled at 8.0 ± 0.2	Scrubber inlet pH controlled at 8.0 ± 0.2
Run Philosophy	Intended 2 wks. To observe sulfite oxidation and degree of sulfate saturation with lime add'n to downcomer, minimum slurry rate to venturi, sealed EHT purged with N ₂ gas, and 8% solids recirculated.	Intended long-term. Control at higher pH in attempt to reduce sulfite oxidation and thereby degree of sulfate saturation. Wash mist eliminator with water only.	Intended short-term. Mist eliminator washed continuously with raw water only at rate greater than available makeup water.	Int'd 2 wks, 12 min res time. Venturi in service. EHT sealed with Freon. Mist elim. on 4 hr wash cycle. Observe effects of lime add'n to downcomer and sealed EHT (compare with Run 001-1A).
Results	Degree of sulfate saturation was about 130%. Solids from outlet duct fell to top of mist eliminator. Run was terminated due to heavy scale (500 mls avg.) and solids buildup on mist eliminator.	Run was terminated due to scale formation (up to 150 mls) on top mist eliminator vanes.	Run was terminated due to scale formation (50 mls avg.) on top mist eliminator vanes.	Run terminated due to slight increase in mist eliminator ΔP. Inspection revealed scale buildup (88 mls avg.) on the mist eliminator top vanes.

Table 3 (continued)

SUMMARY OF LIME RELIABILITY TESTS
ON VENTURI/SPRAY TOWER SYSTEM

Run No.	100-1A
Start of Run Date	9/20/74
End of Run Date	10/2/74
On Stream Hours	270
Gas Rate, acfm @ 140°F	25,000
Spray Tower Gas Vel, fps @ 120°F	5
Venturi Spray Tower Liquid Rate, gpm	100-1200
Spray Tower L/G, gal/mcf	5
Percent Solids Recirculated	7,000/0
Effluent Residence Time, min	24
Solids Disposal System	Clarifier & Filter
Stoichiometric Ratio, moles Ca added/mole SO ₂ absorbed	1.07-1.20
Avg % Limestone Utilization, Total mole SO ₂ abs./mole Ca added	83
Inlet SO ₂ Concentration, ppm	2200-3000
Percent SO ₂ Removed	80-90
Scrubber Inlet pH Range	7.5-8.5
Scrubber Outlet pH Range	4.5-5.4
Percent Sulfur Oxidized	12-19
Leak Closure, % Solids Dischg.	47-52
Calculated % Sulfate Saturation in Scrubber Inlet Liquid @ 140°F	110
Dissolved Solids, ppm	8,000-10,000
Total ΔP Range, Excluding Demister, in. H ₂ O	11.0-11.5
Venturi ΔP, in. H ₂ O	5
Demister ΔP, in. H ₂ O	0.80-0.85
Abatement	Time allowed to 20 gpm with makeup water and added to scrubber discharge.
Mist Eliminator	Bottom washed with makeup at 2.5 gpm/hr for 30 min every 4 hrs. Simultaneous 1.5 g wash with remaining makeup at 1 gpm/hr through a single nozzle covering about 14 in ² . Total makeup = 4 gpm x g.
Scrubber Internals	All nozzles on 4 headers sprayed downward. 1 noz- zle header on top, 3 headers, 4 nozzles on bottom header.
System Changes Before Start of Run	Mist eliminator and outlet duct cleaned. A single nozzle in- stalled to provide 1.5 g wash for low section of mist eliminator and several holes drilled in the top areas of a second section.
Method of Control	Scrubber inlet pH controlled at 8.0 ± 0.2
Run Philosophy	Operated 2 wks. To observe the effect of mist eliminator tip wash for low section and the effect of increased resi- dence time on sulfate saturation.
Results	Run terminated as planned. Sulfate saturation reduced to 110%. Mist eliminator top panels clean where tip washed.

Table 4

AVERAGE SCRUBBER INLET LIQUOR COMPOSITIONS AND CALCULATED SULFATE SATURATIONS FOR VENTURI/SPRAY TOWER LIME RELIABILITY RUNS

Run No.	Percent Solids Recirculated	Effluent Residence Time, min.	Percent Solids Discharged	Percent Sulfur Oxidized	Scrubber Inlet pH Range	Inlet Liquor Species Concentrations, mg/l (ppm)						Calculated Percent Sulfate Saturation at 50°C (d)		
						Ca ⁺⁺	Mg ⁺⁺	Na ⁺	K ⁺	SC ₂ ⁻	SC ₄ ⁻	CC ₂ ⁻	Cl ⁻	Total
0-01-1A ^(a)	7.9	12	20-25	10-30	7.4-8.5	1,000	140	60	80	60	2,000	20	2100	4,400
0-01-1A ^(b)	7.9	12	42-52	10-30	7.5-8.5	3100	200	80	230	60	2,800	25	4,500	10,500
0-03-1A	13.5-16.0	12	46-54	12-22	7.8-8.2	2880	190	90	330	60	1,700	5	4,770	9,490
0-04-1A	7.5-9.0	17	50-60	8-30	7.7-8.4	3,200	400	80	275	50	2,000	60	5,700	11,700
0-05-1A	8.8-9.2	17	48-52	12-28	8.8-9.2	2,200	50	60	180	50	1,450	5	4,070	7,070
0-08-1A	7.5-9.0	12	48-58	13-29	7.9-8.4	2,220	320	60	210	45	1,920	10	3,000	8,700
0-09-1A	8.0	24	46-52	15-25	7.5-8.5	2,800	310	60	120	40	1,440	20	4,260	8,870

Note: Average concentrations are for steady-state operating periods.
 Solids disposal system: Clarifier and filter.

(a) Solids disposal system: Clarifier

(b) Solids disposal system: Clarifier and filter

(c) The values of calculated sulfate saturation are for an average inlet gas SO_2 concentration of 2800 ppm.

(d) Calculated Sulfate Saturation (activity Ca^{++} x activity SC_4^{--}) x solubility SC_4^{--} (solubility product at 50°C). A solubility product for $CaSO_4 \cdot 2H_2O$ of 2.2×10^{-7} was used (Radian Corporation, "A Theoretical Description of the Limestone-Injection Wet Scrubbing Process," NAPCA Report, June 9, 1970).

Reproduced from
best available copy.

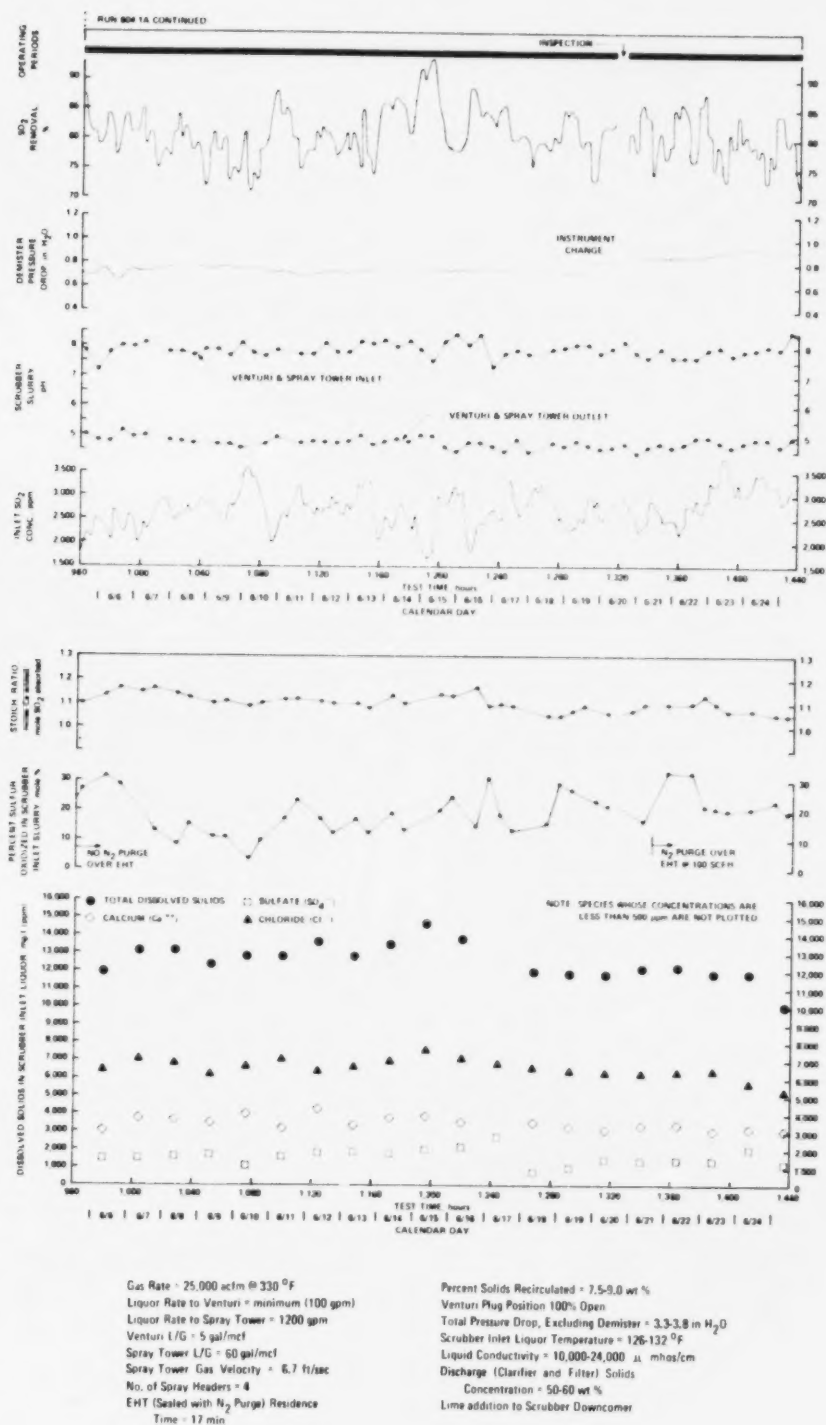


Figure 8. Operating data for Venturi/Spray Tower Run 604-1A

4.1.1 Venturi/Spray Tower Run 601-1A

On October 9, 1973, the initial long-term reliability test (Run 601-1A) was begun on the venturi/spray tower system. The major test conditions selected were (see Table 3):

Spray Tower Gas Velocity	6.7 ft/sec
Venturi Liquid-to-Gas Ratio	30 gal/mcf
Spray Tower Liquid-to-Gas Ratio	60 gal/mcf
Percent Solids Recirculated	8
Effluent Residence Time	12 min. *
Scrubber Inlet Slurry pH (controlled)	8

The pH control level for the scrubber inlet slurry was chosen based on results of lime testing at the EPA pilot facility in Research Triangle Park, which indicated reasonable lime utilization and SO₂ removal at that level.

During the entire test, the chevron mist eliminator was washed on the underside both continuously (at a rate of 0.8 gpm/ft²) and intermittently (at a rate of 1 gpm/ft²) with the available raw water makeup plus clarified liquor. Also, the mist eliminator was washed on the topside once per week with fresh water for 5 minute durations during the last four weeks of the test.

Throughout the initial 666 hour portion of the run the clarifier was used as the final dewatering device and the average percent solids discharged

* Twelve minutes was the minimum residence time obtainable in the effluent hold tank for this run.

was 23 percent, which resulted in slightly open liquor loop operation.* An inspection after 666 hours on November 7 showed that the mist eliminator bottom vanes were clean while the top vanes were about 5 percent plugged with soft solids. The scrubbers were generally clean, with only a thin layer of scale (20 mils) on the upper half of the spray tower.

In order to operate under closed liquor loop conditions, the test was continued with the clarifier plus vacuum rotary drum filter (or centrifuge) in series. Problems with the filter cloth and mechanical difficulties with the centrifuge, however, resulted in intermittent operation of these pieces of equipment from November 7 to December 15. For the final 575 hours of testing (from December 15 through January 8), the filter operated satisfactorily and the average percent solids discharged increased from 23 to 47 percent and the total dissolved solids increased from about 6500 to 10,500 ppm (see Table 4).

Run 601-1A was terminated after a total of 2153 hours (3 months) of operation, due to rapidly increasing pressure drop across the mist eliminator. An inspection showed that the top of the mist eliminator was covered with solids that fell from the duct-work above, plugging about 80 percent of the surface. In addition, about 1/8 to 1/2 inch thick scale formed in the middle and top vanes and about 5 percent of the bottom (inlet) vanes contained heavy (1/4 to 3/4 inch) scale. It is hypothesized that the heavy accumulation of solids in the outlet duct occurred when slurry droplet entrainment through the chevron mist eliminator wetted the duct walls during frequent reheater flame-outs.

* The venturi/spray tower system clarifier is undersized.

Heavy scale had accumulated on the wall of the spray tower below the bottom header while the middle section of the vessel was only slightly scaled. The venturi wall above the throat was clean, but scattered scale covered the plug. A light scale covered the venturi wall between the throat and flooded elbow and the entire flooded elbow was covered with heavy scale. Also, the recirculating piping was scaled, with heaviest scale deposits in the pump suction areas. Scale also formed on the impeller eyes and tips of all the recirculating slurry pumps and on the lining of the casing of some of the pumps. It should be noted that scale formation in the scrubbers, circulating slurry piping and pumps did not prevent continual operation of the system or necessitate termination of the run.

The formation of most of the scale on the scrubber walls and piping occurred between December 15 and January 8, when the system was operating under closed liquor loop conditions. The calculated average sulfate saturation of the scrubber inlet liquor during this period was 180 percent. This is contrasted with a calculated sulfate saturation of 150 percent for the initial 666 hours of testing, during open liquor loop operation (see Table 4).

An analysis of the test during the final 575 hours of closed liquor loop operation shows that the calculated inlet liquor sulfate saturations are strong functions of inlet gas SO_2 concentration (or SO_2 absorption rate). For example, during test periods where the inlet gas SO_2 concentrations averaged 3200 and 2500 ppm, the calculated sulfate saturations averaged 225 and 140 percent, respectively.

4.1.2 Venturi/Spray Tower Run 002-1A

In preparation for the second lime reliability run, the system was partially cleaned chemically (Na_2CO_3 /sugar/limestone/fly ash solution), followed by mechanical cleaning. All the stainless steel spray nozzles were replaced with identical, full cone, stellite-tipped Bete nozzles (ST 48-FCN) and the original Hauck reheater, which was troublesome throughout the previous run, was replaced with an external combustion reheater (by Bloom Engineering Co.).

Results from the EPA pilot facility in Research Triangle Park had indicated that sealing the effluent hold tank could reduce the sulfite oxidation and, thereby, the degree of sulfate saturation of the scrubbing slurry. In an attempt to duplicate this mode of operation, the effluent hold tank on the venturi/spray tower system was sealed prior to the start of Run 002-1A.

Run 002-1A was started on March 15, 1974. The mist eliminator was washed on the underside with the available raw water makeup and clarified liquor (in an approximate one to seven ratio) at a rate of 1 gpm/ft², in an intermittent operation (4 minutes on/1 minute off cycle).

After 393 operating hours the run was terminated, when it became apparent that sealing the effluent hold tank in this facility was not effective in reducing the degree of sulfate saturation of the scrubber inlet liquor. Although the sulfite oxidation had dropped somewhat (from about 20 to 17 percent), the calculated sulfate saturation at the termination of the test was 190 percent.

An inspection showed that a uniform (about 1/8 inch) scale covered most of the top and middle vanes of the mist eliminator. The bottom vanes were relatively clean. The walls of the spray tower were covered with a thin scale. The wall of the venturi below the plug was generally clean and no scale was deposited in the flooded elbow.

4.1.3 Venturi/Spray Tower Run 603-1A

Run 603-1A was started on April 2, 1974, after the mist eliminator was cleaned. The solids content of the recirculating slurry was increased from 8 to 15 percent in order to reduce the degree of scrubber liquor sulfate saturation. The mist eliminator was washed on the underside with the available raw water makeup and clarified liquor (in an approximate one to four ratio) at a rate of 1 gpm/ft², in an intermittent operation (1 1/2 minutes on/1 1/2 minutes off).

After 395 hours of operation the run was terminated, due to increasing pressure drop across the mist eliminator (from 0.55 inches H₂O at the start of the run to 0.70 inches H₂O). The mist eliminator had 50-70 mils of scale on the outlet edge and soft solids along the inlet edge of the top vanes. The middle and bottom vanes had light deposits of scale and soft solids. The venturi scrubber contained about the same amount of scale as before the start of the run, while there appeared to be slightly less scale on the spray tower walls and in the suction piping of the slurry pumps. The calculated sulfate saturation of the scrubber inlet liquor was 135 percent. This was a significant drop from the previous run's value of 180 percent saturation at 8 percent solids recirculated.

4.1.4 Venturi/Spray Tower Run 604-1A

In order to decrease the quantity of soft solids buildup on the mist eliminator which had caused termination of the previous test, it was decided to drop back the percent solids recirculated from 15 to 8 percent for this run. To reduce the scrubber slurry sulfate saturation from the expected level of 180 percent (see Run 601-1A) (1) lime was added to the scrubber downcomer (vs. the effluent hold tank), (2) the effluent hold tank was purged (blanketed) with N_2 , and (3) the venturi scrubber was used only for gas cooling (venturi plug 100 percent open with 100 gpm liquor flow rate).

It was theorized that the addition of lime to the downcomer quickly increases the pH of the scrubber outlet liquor, thereby precipitating calcium sulfite before the sulfite can be oxidized to sulfate in the liquid phase (the solubility of calcium sulfite is lower at higher pH). As mentioned previously, a reduction of liquor phase sulfite oxidation results in a reduction in sulfate saturation. It was further theorized that reducing the liquor flow to the venturi with a minimum flue gas pressure drop would minimize liquid atomization, oxygen absorption and, consequently, oxidation.

Run 604-1A was begun on April 26, 1974, after the mist eliminator was cleaned. The mist eliminator was washed on the underside with the available raw water makeup and clarified liquor (at an approximate one to seven ratio) at a rate of 1 gpm/ft², in an intermittent operation (3 1/2 minutes on/1 1/2 minutes off). The effluent residence time was increased to 17 minutes, since this was the minimum time obtainable

at the reduced total liquor rate of 1300 gpm (the venturi liquor rate had been reduced from 600 to 100 gpm).

The run was terminated after 1828 on-stream hours due to (1) increasing pressure drop across the mist eliminator (from 0.62 inches H_2O at start of run to 1.75 inches H_2O), and (2) gas flow control problems caused by solids accumulation on the ID fan inlet dampers. The top of the mist eliminator was 75 percent plugged with 150-200 pounds of loose solids which had fallen from the outlet duct. There was about 1/2 inch of scale on the top vanes and 1/4 inch of scale on the bottom vanes. Up to 2 inches of solids had accumulated in the outlet gas duct and six of the 14 damper blades were covered with 1-2 inches of solids.

The sulfite oxidation averaged from about 14 percent with N_2 purge over the effluent hold tank to about 19 percent without N_2 purge (from May 29 to June 5).^{*} The calculated average sulfate saturation of the scrubber inlet liquor, however, was 130 percent. This was a significant drop from the saturation level of 180 percent during the closed liquor loop portion of Run 601-1A.

As in Run 601-1A, an analysis of the data showed that the calculated inlet liquor sulfate saturation was a strong function of the SO_2 absorption rate. For example, during test periods where the inlet gas

^{*} Oxygen concentrations in the effluent hold tank in the vicinity of the gas/liquid interface ranged from 15 to 20 percent, as compared with 21 percent for air. This suggests that sealing the tank and N_2 purging were not effective in reducing oxygen absorption in the effluent hold tank.

concentrations averaged 2700 and 2900 ppm, the calculated sulfate saturations averaged 120 and 150 percent, respectively.

Operating data for Run 604-1A, from June 6 to June 24, 1974, has been graphically presented in Figure 8.

4.1.5 Venturi/Spray Tower Run 605-1A

Run 605-1A was started on July 31, 1974, after the mist eliminator was cleaned. The scrubber inlet liquor pH was raised from 8 to 9, in order to observe the effect of pH on sulfate saturation. The mist eliminator was washed intermittently with raw water makeup at a rate of 0.4 gpm/ft² on a 1 minute on/4 minute off cycle. As before, lime was added to the downcomer and the venturi liquor rate was set at 100 gpm with the venturi plug 100 percent open.

An inspection after 141 operating hours showed relatively heavy (up to 150 mils) scale on the top mist eliminator vanes and the run was terminated. As with the previously observed patterns, there was a moderate buildup of scale on the middle vanes (50 mils) and only a light dust film on the bottom vanes. No additional scale had formed in the spray tower during the run. The calculated average percent sulfate saturation of the scrubber inlet liquor was 115 percent. This can be compared with a level of 130 percent for Run 604-1A, which had a controlled scrubber inlet liquor pH of 8.0 (see Table 4).

4.1.6 Venturi/Spray Tower Run 606-1A

Run 606-1A was started on August 7, 1974, after the mist eliminator had been cleaned. The objective of the run, which was intended to be

of short duration, was to investigate scale formation on the mist eliminator vanes with continuous washing on the mist eliminator underside with raw water at a rate of 0.3 gpm/ft². This raw water rate (15 gpm) is approximately three times the quantity normally used during closed liquor loop operation, and, therefore, the test was made under open liquor loop conditions (about 21 percent solids discharged). The scrubber inlet liquor pH was reduced from 9 to 8 for this test.

The run was terminated after 170 on-stream hours. About 95 percent of the trailing (top) surface of the mist eliminator vanes was covered with 30-70 mils of scale. As usual, there was light soft solids accumulation on the middle vanes and the bottom vanes were relatively clean. It should be noted that the scale growth rate on the top vanes (~50 mils/week) was about the same for this test as for all of the previous tests, in which mixture of clarified liquor plus raw water were used for underside washing.

4.1.7 Venturi/Spray Tower Run 608-1A

Run 608-1A was started on August 21, 1974, after the mist eliminator had been cleaned. The main objective of the test was to investigate scale formation on the mist eliminator vanes with intermittent underside washing at high pressure (60 psig), using the available raw water makeup at a rate of 3 gpm/ft² for about 9 minutes every 4 hours. Also, the venturi scrubber was put back in service (at 600 gpm and 9 inches H₂O pressure drop), in order to investigate the effect of venturi operation (Runs 608-1A vs. 604-1A) and lime addition in the downcomer (Runs 608-1A vs 601-1A) on scrubber liquor sulfate saturation.

The run was terminated after 610 hours of operation due to increasing pressure drop across the mist eliminator (from 0.75 inches H₂O at start of run to 0.95 inches H₂O). About 80 percent of the trailing surface of the top vanes was covered with scale, averaging 88 mils in thickness and resulting in approximately 3 percent pluggage of the free flow area. The middle and bottom vanes were, as usual, essentially clean (3 mils of dust coverage). Note that, the average scale growth rate on the top vanes for this test (24 mils/week) was essentially half the average scale growth of the previous four tests.

The condition of the scale on the walls of the spray tower remained essentially unchanged, but the spiral tips of most of the spray nozzles were covered with 1/4-1/2 inch scale (scale "whiskers").

The calculated average sulfate saturation of the scrubber inlet liquor was 130 percent (see Table 4). This is the same as the calculated saturation for Run 604-1A and about 50 percent lower than the calculated saturation for Run 601-1A. Therefore, it may be concluded that, for these run conditions, venturi operation has little effect on sulfate saturation and the addition of lime to the downcomer (vs. lime to the effluent hold tank) significantly reduces the sulfate saturation of the process slurry.

4.1.8 Venturi/Spray Tower Run 609-1A

Run 609-1A was started on September 20, 1974, after the mist eliminator had been cleaned. The primary objective of the test was to evaluate the effect of topside mist eliminator washing on the formation of scale

on the top mist eliminator vanes. The entire underside of the mist eliminator and a 14 ft² area on the topside was washed at high pressure (60 psig) with makeup water at a rate of 2.7 gpm/ft² for the underside and 1.0 gpm/ft² for the topside for about 8 minutes every 4 hours. Only a small section of the topside was washed because it was felt that liquid carryover from top sprays could possibly cause reheater overloading and fan problems. Also the effluent residence time was increased from 12 to 24 minutes, for this test, in order to determine the effect of residence time on scrubber liquor sulfate saturation (Run 609-1A vs. 608-1A).

Run 609-1A was intended to be of short duration and was terminated after 278 hours of operation. The topside spray had been successful in drastically reducing the scale buildup on the top mist eliminator vanes. The washed area was essentially clean, with less than 1 mil of solids accumulation, compared with an average of 40 mils scale buildup on the rest of the topside surfaces.

The calculated average sulfate saturation of the scrubber inlet liquor was 110 percent for the test (see Table 4). This value is about 20 percent lower than the calculated saturation for Run 608-1A. It may be concluded, therefore, that the effect of effluent residence time on sulfate saturation is similar for lime systems as for limestone systems (see Figure 6).

These problems could be reduced by a sequential sectional wash of the topside of the mist eliminator or by using a second mist eliminator downstream of the topside sprays.

4.2 CONCLUSIONS

4.2.1 Scrubber Internals Operability

The lime and limestone reliability tests have shown that scrubber internals can be kept relatively free of scale if the sulfate (gypsum) saturation of the scrubber liquor is kept below about 135 percent. As with limestone, this can be accomplished with increased percent solids recirculated (Run 603-1A vs. 601-1A) and/or with increased effluent residence time (Run 609-1A vs. 608-1A). Moreover, the lime tests have shown that adding the lime to the scrubber downcomer (which corresponds to a small residence time tank in series with the larger effluent hold tank) can substantially reduce the sulfate saturation (Run 608-1A vs. 601-1A), allowing for operation at reduced percent solids and/or effluent residence time. Also, operating at higher scrubber inlet liquor pH appears to reduce the scrubber liquor sulfate saturation (Run 605-1A vs. 604-1A).

The lime tests have also shown that the sulfate saturation of the scrubber inlet liquor is a strong function of the inlet gas SO_2 concentration (SO_2 absorption rate). The data has indicated that a 100 ppm increase in SO_2 inlet concentration corresponds, roughly, to a 13 percent increase in sulfate saturation, at the run conditions tested.

It should be noted that the 28 full cone stellite-tipped Bete nozzles in the spray tower have shown no signs of measurable erosion after approximately 4000 hours in slurry service.

4.2.2 Mist Eliminator Reliability

The most significant reliability problem encountered, to date, during the venturi/spray tower lime reliability tests has been associated with scaling and/or plugging of the topmost vanes of the chevron mist eliminator (see Figures 1 and 3).

Runs 601-1A through 608-1A showed that, the scale formation rate on the top mist eliminator vanes (25-50 mils/week) was relatively unaffected by the wash cycle, wash rate or quality (sulfate saturation) of the wash liquor. Run 608-1A showed that intermittent washing with high pressure (60 psig) raw water at a rate of 3 gpm/ft² for 9 minutes every 4 hours, gave results that were at least as good as results with continuous washing with relatively low pressure raw water at 0.3 gpm/ft² (Run 606-1A). Intermittent washing can be especially important if the availability of raw water makeup is restricted (e.g., at high percent solids discharged and/or high pump seal water usage).

Run 609-1A showed that intermittent topside and bottomside washing with high pressure raw water would most likely eliminate the scale accumulation problem on the top vanes of the existing mist eliminator (see Figure 3) and allow for long-term reliability of the system at the run conditions tested (see Table 3). However, the problems caused by entrainment of topside wash water will have to be evaluated.

The reliability of the intermittent topside and bottomside washing with raw water has been further substantiated in a recently completed 253 hour test (Run 610-1A). The scrubber and mist elimination systems

were not cleaned before the test. After a total of 531 operating hours (Runs 609-1A and 610-1A), the washed area of the top mist eliminator vanes was still clean, compared with an average of 34 mils scale buildup on the rest of the topside vane surfaces.

It is planned to test a sloped (cone-shaped) closed-vane chevron mist eliminator in the spray tower in the near future. It is expected that the sloped mist eliminator has better draining characteristics than a horizontal mist eliminator.

Section 5

PARTICULATE REMOVAL TEST RESULTS

The overall particulate removal efficiencies measured during the reliability testing on the TCA and venturi/spray tower systems are shown in Tables 5 and 6, respectively. For these tests, a modified EPA particulate train (manufactured by Aerotherm/Acurex Corporation) was used to measure mass loading at the scrubber inlets and outlets. As can be seen from the tables, the measured removal efficiencies are all greater than 99 percent for the tests conducted.

In May 1973, the EPA measured overall particulate removal efficiency and particulate removal efficiency as a function of particle size for the Shawnee TCA scrubber during limestone reliability verification testing (see Section 3.3 in Reference 2). The EPA measurements for overall efficiencies are in good agreement with the results shown in Table 5. The mass mean diameter of the inlet solids, as determined by EPA, was approximately 23 microns.

The particulate removal efficiencies shown in Tables 5 and 6 appear to be higher than the efficiencies predicted from "impaction theory." These improved efficiencies could be due to condensation of water vapor in the flue gas on the solid particles. In order to verify the

Table 5

OVERALL PARTICULATE REMOVAL IN TCA SCRUBBER FOR
LIMESTONE RELIABILITY TESTS

Run No.	Date	Gas Rate, acfm @ 300°F	Liquor Rate, gpm	Pressure Drop, (a) in. H ₂ O	Grain Loading, grains/scf		Percent Removal
					Inlet	Outlet	
531-2A	5/15	20,500	1200	6.7	2.82	0.029	99.0
531-2A	6/12	20,500	1200	10.0	3.47	0.010	99.7
532-2A	7/24	20,500	1700	6.7	3.33	0.024	99.3
535-2A	9/25	20,500	1200	6.2	2.88	0.016 ^(b)	99.4

(a) Including mist eliminator and Koch tray.

(b) Outlet sample taken on 9/27, two days after inlet sample.

Table 6
OVERALL PARTICULATE REMOVAL IN VENTURI/SPRAY TOWER
FOR LIME RELIABILITY TESTS

Run No.	Date	Gas Rate, acfm @ 330°F	Liquor Rate, gpm		Pressure Drop, in. H ₂ O		Grain Loading, grains/scf		Percent Removal
			Venturi	Spray Tower	Venturi	Spray Tower (a)	Inlet	Outlet	
604-1A	6/26	25,000	100	1200	1.9	3.0	2.52	0.024	99.1
608-1A	9/11	25,000	600	1200	9.0	2.2	2.28	0.023	99.0
610-1A	10/10	25,000	600	1200	9.0	3.2	2.72	0.021	99.2

(a) Including mist eliminator.

particulate removal results obtained, to date, and identify the causes for the observed removals. EPA has planned a new test series for the determination of size and overall removal efficiencies in the TCA and venturi/spray tower systems. The test series is to be conducted by an independent party in the near future.

Section 6

REFERENCES

1. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Phase II and III Final Report, EPA Report, December 1974 (to be issued).
2. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Limestone Wet Scrubbing Test Results, EPA Report 650/2-74-010, January 1974.
3. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Sodium Carbonate and Limestone Test Results, EPA Report 650/2-73-013, August 1973.
4. Bechtel Corporation, Test Manual for Advanced Test Program, EPA Report, September 1974.
5. P. S. Lowell, "Use of Chemical Analysis and Solution Equilibria in Predicting Sulfate/Sulfite Scaling Potential," presented at Second International Lime/Limestone Wet Scrubbing Symposium, New Orleans, Louisiana, November 8-12, 1971.
6. R. H. Borgwardt, "EPA/RTP Pilot Studies Related to Unsaturated Operation of Lime and Limestone Scrubbers," presented at Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.
7. Radian Corporation, A Theoretical Description of the Limestone-Injection Wet Scrubbing Process, NAPCA (APTIC No. 22709 and 25446) Report, June 9, 1970.

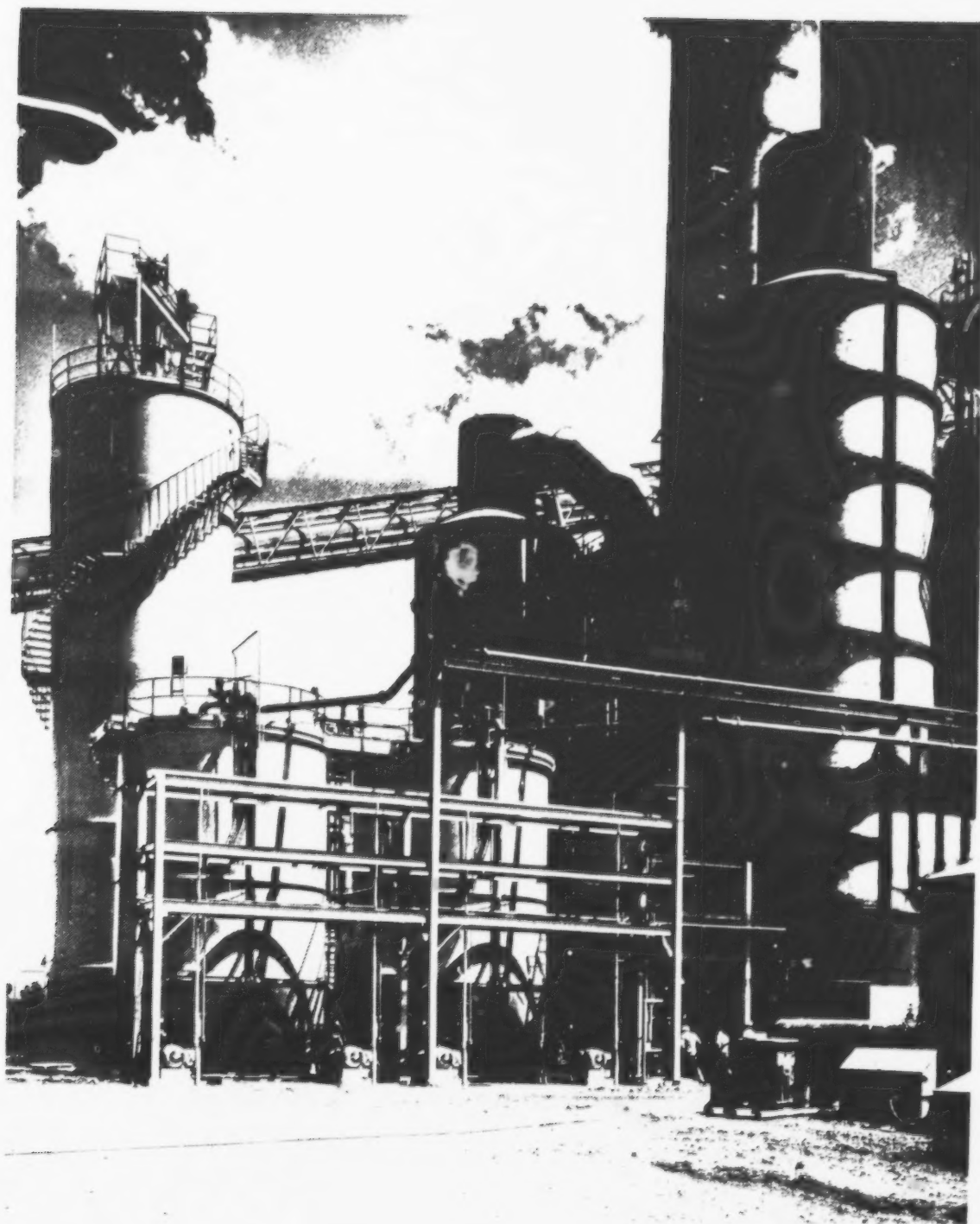
OPERATIONAL STATUS AND PERFORMANCE
OF THE
ARIZONA PUBLIC SERVICE COMPANY
FLUE GAS DESULFURIZATION SYSTEM
AT THE CHOLLA STATION

By

Lyman K. Mundth

Arizona Public Service Company

November 4, 1974



OPERATIONAL STATUS AND PERFORMANCE
OF THE
ARIZONA PUBLIC SERVICE COMPANY
FLUE GAS DESULFURIZATION SYSTEM
AT THE CHOLLA STATION

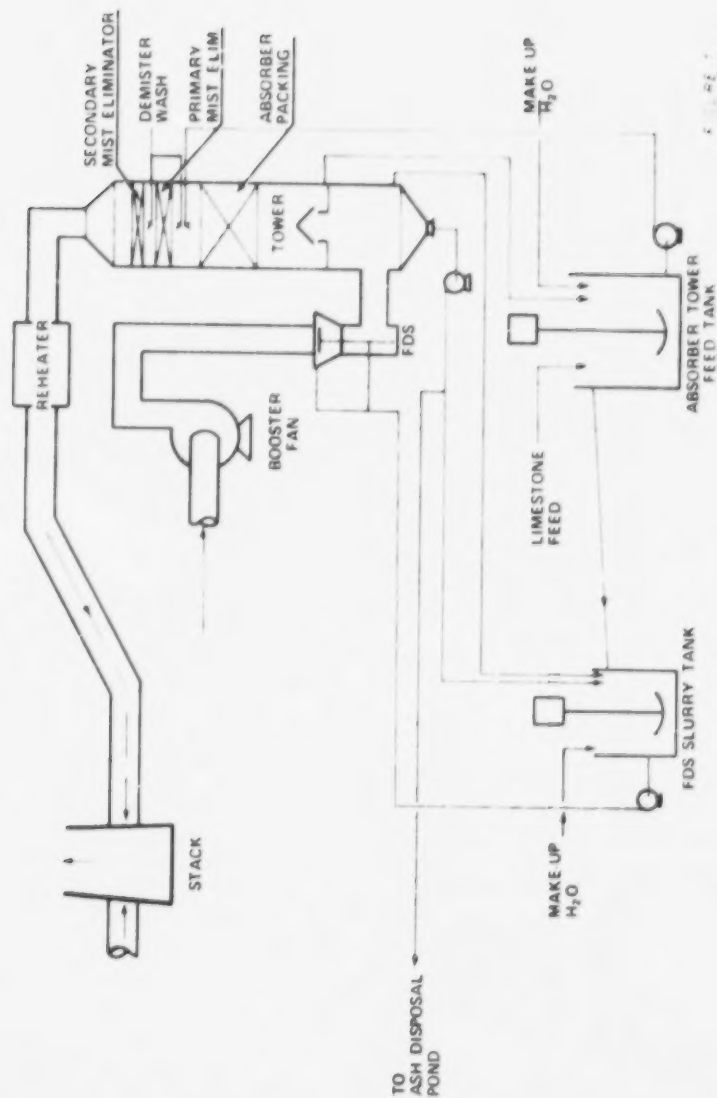
My name is Lyman Mundth and I am Vice President of Power Production and Operations for Arizona Public Service Company. My discussion will cover our experiences with wet scrubber systems at our Cholla Power Plant near Joseph City, Arizona. The Cholla Plant currently has a single coal fired unit rated at 115 MW which has been operating since 1962. The unit burns coal from the Gallup, N.M. area and this coal contains approximately 10 percent ash and .5 percent sulfur.

We operate this unit base loaded and have historically had trouble free operation, with a capacity factor of near 90 percent. The original capital cost of the installation in 1962 was approximately \$20 million. Mechanical collectors designed to remove 80 percent of the fly ash provided the emissions abatement on that installation. We recently retrofitted the unit with a Research-Cottrell sulfur dioxide/fly ash removal process and a by-pass system at a capital cost of \$6.3 million.

Construction of the Research-Cottrell system was completed in December, 1973, one year behind the scheduled date of December 3, 1972. The system, consisting of two parallel gas cleaning trains, began commercial operation in mid-December, 1973, and has operated to the present with only short outages. Each gas cleaning train consists of a Flooded Disc Scrubber, a wetted film absorber, mist eliminators, and reheater. The limestone and fly ash slurry handling systems are common to both trains.

For those of you who picked up a copy of my talk, you will notice that Figure 1 illustrates a single train flow diagram. Flue gas containing SO_2 and remaining fly ash flows from the outlet of the mechanical collectors to the inlet of the wet scrubber system. A forced draft booster fan forces the gas into the flooded disc scrubber where the fly ash and a portion of the SO_2 are removed by a recirculating limestone/fly ash slurry. This slurry enters into the flooded disc scrubber tangentially above the throat and at the disc. Particulate

CHOLLA UNIT 1 WET SCRUBBER SYSTEM



emissions have been tested and have been found to average 0.026 pounds per million BTU, well within state and federal emissions regulations.

About 70 to 85 percent of the sulfur dioxide passes through the flooded disc scrubber into a cyclonic mist eliminator where the captured ash and entrained slurry are removed. The flue gas then passes into a packed tower which has 2 to 5 feet of Munters packing, a wetted film packing. A limestone slurry recirculates through the absorber packing and absorbs SO_2 . The flue gas and slurry are contacted countercurrently. To minimize scaling, a high liquid to gas ratio and a moderate reaction products concentration are maintained. Total SO_2 removal in the train containing packing is expected to be around 80 percent. However, no comprehensive tests on SO_2 removal have been conducted to verify this removal level.

At this point, I should explain that, due to the uncertainties involved in SO_2 scrubbers, only one of the trains actually contains packing material.

A two-stage mist eliminator follows the absorber. The first stage is an impingement vane type which is flushed intermittently with water to avoid both supersaturation and encrustation on the surfaces. The second stage is also an impingement vane type.

Before entering the stack a steam coil reheater constructed of 316 stainless steel heats the cleaned flue gas approximately 400 F.

The slurry handling system consists of a limestone storage silo and absorption tower slurry, flooded disc and limestone mix tanks.

The system is not equipped with any special sludge and fly ash disposal equipment. Sludge and fly ash are pumped via temporary surge tanks to pre-existing ash disposal ponds. However, we do have plans to provide sludge disposal capacity for this system in the new units currently under construction at the plant.

Although APS has operated the scrubbers since completion of construction in December of 1973, many startup problems have prevented us from final acceptance of these units. We had scheduled acceptance tests in February of 1974. We have not run these tests due to modifications required on the system. Some modifications have been relatively minor while others have required unit outages to accomplish. We expect acceptance tests to be completed before the end of 1974.

Major problems encountered to date can be classified into three categories. They are scaling and plugging, corrosion and mechanical. We have found scaling in the scrubber vessel, in piping and in the mist eliminator. Plugging occurs in the nozzles to the flooded disc and the mist eliminator. Corrosion has been experienced in the reheater and breeching expansion joints. We have encountered mechanical problems in the flooded disc control system, by-pass damper and the reheater.

I would like to discuss some of these problems briefly. First, let's consider our mechanical problems. On initial startup of the unit, vibration in the reheat sections limited operation of the scrubbers to such an extent that commercial operation had to be delayed until a solution could be found. Research-Cottrell hired an outside consultant to look at the vibration problem and to determine a solution. The installation of baffles was recommended because Research-Cottrell's consultant felt that improper flue gas distribution was causing the vibration problem. This fix initially appeared to have solved the problem. Recently, however, due to other fixes, mainly on the leaking by-pass control dampers, vibration is re-occurring at full load, but is reduced considerably in frequency. The reheater vibration is not considered to be a major problem at this time, but we are working toward a final solution.

We have also had problems with the by-pass damper operation. The scrubber system is equipped with a by-pass system so that the boiler may continue to operate should any major failures occur in the scrubber system. Leakage due to sluggish operation and temperature caused distortions of the by-pass dampers have reduced the effectiveness of the scrubber system. Small amounts of untreated flue gas pass through the closed damper and by-pass the scrubber system. Although I have previously mentioned that the system has been tested and found to control particulate to a level of 0.026 pounds per million BTU, we are unable to consistently achieve this level due to improper operation of the by-pass damper. However, we feel the scrubbers themselves have been attaining their designed particulate removal efficiency.

Since the entire installation was a retrofit, the damper system was installed in the existing breeching well in advance of actual scrubber operation. Fly ash build up occurred and prolonged exposure to high temperatures caused excessive expansion and distortion. The boiler had to be shut down to trim, clean and align damper louvers. This did not completely solve the problem. Research-Cottrell then reduced the pressure drop required in the scrubber system in order to

create a recirculating effect such that a small amount of treated flue gas would actually flow backward through the damper and thereby eliminate leakage of untreated flue gas. Although no prolonged operation or testing has occurred, we hope this solves the problem.

A third mechanical problem involves the position control of the flooded disc. The Research-Cottrell Flooded Disc Scrubber maintains a constant pressure drop at all loads by controlling the position of the flooded disc. Some minor problems were experienced in this area. Initially the disc kept oscillating between positions. Build-up around the disc shaft caused binding and improper operation. There was also a problem of control points being set too near control limits. Fortunately, this particular problem was easily resolved by Research-Cottrell.

Other mechanical problems which bear mentioning are leaks in the booster fan which have been experienced due to improper welding and erosion of pump impellers and scrubber vessels. Generally, mechanical problems have been shakedown in nature and have not been totally unexpected.

Next I would like to cover corrosion problems. We have had corrosion problems in the reheater due to condensation in the duct work leading to the reheater. Initially the duct work was uninsulated and in cold weather acid condensation occurred. The acid run-off caused necking of tubes at the reheater sheet and subsequently tubes have failed. In order to solve the problem, Research-Cottrell has insulated the breeching to minimize or eliminate the condensation. We are not sure at this point whether this has solved the problem because no cold weather has been experienced since the insulating job.

We have observed pitting in other tube areas, however, we do not know whether this is due to the manufacturing process or to chemical attack. Only one small leak has been observed other than those due to acid run-off and its cause is unknown.

Because of the condensation in the breeching, failures have also been experienced in the expansion joints. We have just recently replaced all of the metal expansion joints with a rubberized fabric type. Very little operating experience is available with the new joints.

Other corrosion problems which have occurred have been due to improper application of protective coatings and were easily repaired.

It is our belief that plugging in parts of the scrubber system has occurred due to excessive scaling caused either by settling or high concentration of solids. Chemical control on the system has been good.

Material has accumulated and partially plugged the first stage demister, but this material has been soft and easily removed by washing with high pressure water. No adverse operating effects have been seen due to excessive plugging, however, Research-Cottrell has re-designed the demister wash system for better coverage and frequency of washing. It now appears from a short operating experience that only minimal buildup occurs.

Piping has scaled in the tangential nozzle area of the flooded disc scrubber which in turn has caused nozzles to plug. The nozzles are designed in such a way that they can be cleaned while the system is in operation. The plugging is due mainly to settling out of solids in the piping to the nozzles and the formation of a scale plug. Research-Cottrell has re-designed this system with increased velocities, however, the problem has not been totally eliminated.

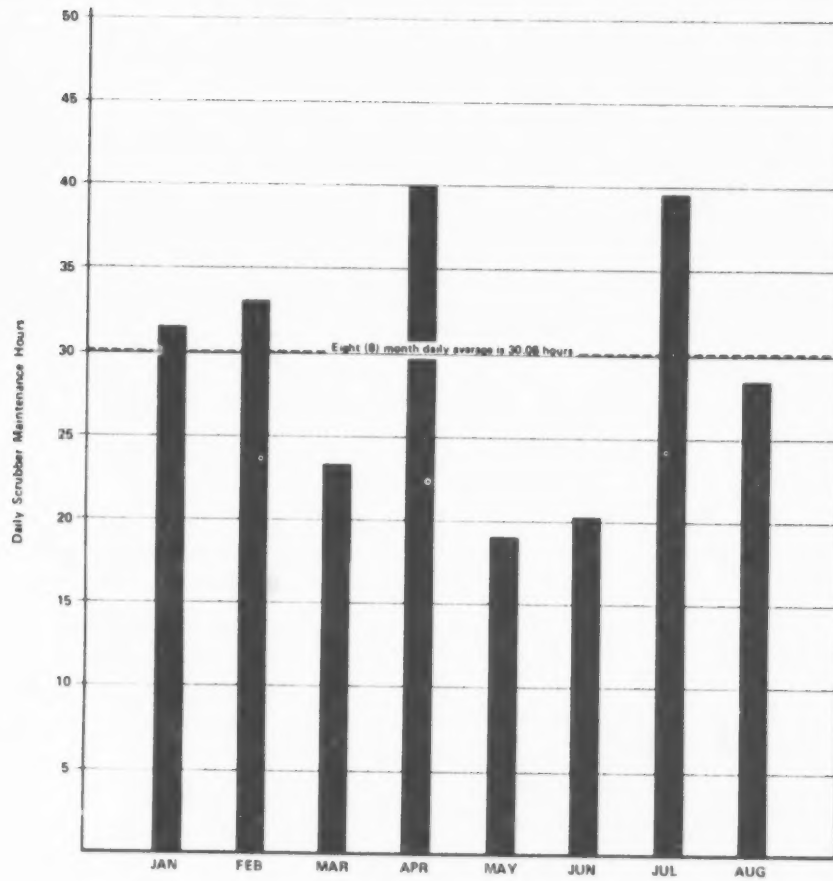
Problems in most of the areas discussed appear to be those which will be experienced in most scrubber systems. These will undoubtedly require a good maintenance effort.

Scrubbers appear to require a great deal more maintenance than other equipment in general use by utilities. We expect that we will have additional maintenance on nozzles, scale removal, valve wear, pump packing and impellers and process lines. Although the system shakedown is continuing, a tremendous effort has already been put forth by both APS and Research-Cottrell during the time that the scrubbers have been in operation. We hope that the same type of effort will not be required in normal operation.

We have tried to separate the shakedown effort from the maintenance effort in order to project our maintenance requirements for the scrubbers. Again, for those of you who have copies, Figure 2 illustrates manhour requirements for experienced scrubber maintenance. The average daily manhour requirement through August of this year has been about 30 hours. This constitutes 37 1/2 percent of the available maintenance manpower at this plant. We expect that it will be necessary to increase our maintenance force by approximately 50 percent to avoid sacrificing other routine and preventive plant maintenance programs and total unit reliability.

CHOLLA UNIT 1 - SCRUBBER MAINTENANCE

Figure 2



In addition, one auxiliary operator per shift is required for scrubber operation. Auxiliaries for the scrubbers include 2.8 MW for electricity and 18,000 pounds per hour of steam for reheat. Limestone is consumed at the rate of 15 tons per day. As indicated previously the system currently does not include waste disposal.

Table 1 is a summary of operating costs for the scrubber at the bus bar. Based on these costs and on our 1973 operating costs, it is estimated that the scrubber will add approximately 15.9 percent to the cost at the bus bar.

TABLE 1
CHOLLA SCRUBBER OPERATING COSTS
(ESTIMATED)

	<u>Mills/KWH</u>
Auxiliaries (Steam and Electrical)	0.2
Operating Labor	0.09
Maintenance Labor	0.09
Parts and Materials	0.07
Limestone	0.15
TOTAL	0.6 mills/KWH
Total Cost 1973 FPC Form 1	3.78 mills/KWH

With regard to system availability, scrubber A which contains packing in the absorber tower has operated for a total of 4,778 hours vs a possible 5,064 hours. Scrubber B has operated a total of 4,401 hours vs a possible 5,064 hours. These numbers do not accurately represent the scrubber availability since a great deal of time has been spent in resolving the problem areas in order to minimize down time. During times when leaks have occurred in the reheater or nozzles have plugged, the unit continued to operate so that individual items of equipment are not necessarily represented by the above figures. No hard availability data has been kept and it is unfair to suggest that one time malfunctions which have been corrected by design changes be included in any assessment of true system availability in the future.

While the scrubbing system at Cholla appears to be successful, it does not necessarily reflect the experiences of others on large utility boilers. On very large boilers, solutions which have worked at Cholla may not work due to increased size and numbers of scrubbers. It is our belief that improvements must be made in the future and possibly new approaches of dealing with sulfur must be developed and

tried in order to successfully handle variations in size, fuel types and environmental constraints.

In summary, we feel the system will ultimately work. However, new developments are needed. The systems must be improved both mechanically and chemically. Disposal of wastes from existing throw-away systems is a serious problem not yet solved. Hopefully, technology will find a way to utilize or eliminate these wastes. Technology should be given adequate time to provide the solutions at a realistic cost in terms of dollars and reliability of generation.

**Wet Scrubber Operating Experience
at
La Cygne Station Unit No. 1**

**Kansas City Power & Light Company
Kansas Gas & Electric Company**

**By C.F. McDaniel
Kansas City Power & Light Co.
Kansas City, Missouri**

**Presented at
EPA Flue Gas Desulfurization Symposium
Sheraton-Biltmore Hotel
Atlanta, Georgia
November 4-7, 1974**

319

Preceding page blank



Description

La Cygne Unit No. 1 is a new generating station located about 55 miles south of downtown Kansas City and east of La Cygne, Kansas, figure 1. The plant is jointly owned by Kansas City Power & Light and Kansas Gas & Electric Company. The plant is manned by KCP&L employees, with output shared equally by the two companies.

Site construction was started in April, 1969, with erection of the Air Quality Control System initiated in mid-1971. Unit No. 1 was nearing completion when the boiler was first fired on December 26, 1972, and construction was considered complete when the unit was declared commercial on June 1, 1973.

An earthfill dam almost 7,000 feet long was constructed to form a 2,600 acre lake for cooling. Flyash and spent slurry from the scrubber are contained in a 160 acre settling pond located east of the cooling water lake.

Coal is delivered to the mine-mouth plant with off-the-road 120-ton trucks from a strip mine operated by Pittsburg & Midway Coal Company. Run of mine size coal, 30 in. x 0 is conveyed from the receiving hoppers at a rate of 2,000 tph to two rotary breakers and reduced to 3 in. x 0 size. The fuel is a low grade bituminous coal that has an as-fired heating value of 9,000-10,000 Btu/lb, 20-30 percent ash, and up to 6 percent sulphur. The coal continues on to be stacked in the storage area with the stacker-reclaimer, or a maximum of 1200 tph may be bypassed to the secondary crushers.

Limestone is obtained from nearby quarries where the rock is reduced to 3/4 in. x 0. It is delivered to the plant in 50-ton off-the-road trucks and is stacked and conveyed with the coal handling equipment.

The boiler is a cyclone-fired, supercritical, once-through, balanced draft Babcock & Wilcox boiler. Its rating is 6,200,000 pounds of steam per hour, 1,010°F, 3,825 psig at the superheat outlet and 1,010°F at the reheater outlet.

The turbine-generator was supplied by Westinghouse and is rated at 870 MW gross output with 5 percent overpressure.

The air quality system, shown schematically in Figure 2 was supplied by Babcock & Wilcox, and consists of seven, one-seventh capacity venturi-absorber scrubber modules, designed to handle the entire flue gas flow of 2,370,000 ACFM. (Figure 3)

Each scrubber module consists of a variable-throat venturi followed by a two-stage countercurrent tray absorber, with the wetted parts fabricated from 316L stainless steel. Seven rubber lined, 7,750-GPM venturi recirculation pumps and seven 10,300-GPM absorber recirculation pumps recirculate a flyash-limestone slurry in the system.

The original plant design called for only one of the seven scrubber modules to be supplied with two stages of absorber internals and absorber recirculation pumps.

Flue gas leaving the scrubber modules is reheated 25 F with a bare-tube steam coil reheater and discharged to a 700-foot stack by six one-sixth capacity induced draft fans which draft both the boiler and scrubber system.

Limestone rock, 3/4 in. x 0 pulverized to a fineness of 95 percent minus 200 mesh in two 100 tph Koppers wet ball mills.

The auxiliary power requirements for the AQC system, 28 MW, are about the same as the normal auxiliaries for the plant, 22 MW. The total auxiliaries result in a design plant net electrical output of 820 MW.

Startup and Operation

The scrubber system was initially placed in service on December 26, 1972, using two venturi scrubber modules during initial firing of the boiler. The turbine was rolled on January 25, 1973, and synchronized on February 23, 1973. A total of 16 days of operation was obtained with the boiler-scrubber during the first 3-month startup period. Sustained unit load was not obtained until the month of April, 1973. The unit was declared commercial on June 1, 1973, only 4 weeks after the scheduled commercial date, and a peak load of 832 MW obtained on June 2, 1973. During this initial trial operating period, many problems were

encountered with the plant, most of which would not be considered abnormal when starting up an entire new power plant facility the size of La Cygne. These problems involved auxiliary boiler casing leaks, boiler tube leaks, furnace slag tap freeze-ups, generator balancing, control system problems, coal bunker difficulties, and auxiliary transformer problems.

Scrubber system problems during the startup period involved mainly completing the construction of scrubber modules and the instrumentation and controls.

One of the major operating problems has been associated with the induced draft fans being designed too close to their critical speed, which resulted in high fan vibrations created by buildup of wet flyash deposits on the rotors. This condition necessitated frequent fan washing and balancing of the rotors. The fan rotors have all been rebuilt with increases in shaft diameter and blade thickness in order to increase the critical speed. The problem of wet flyash deposits on the fans is thought to be the result of a number of conditions which exist in the scrubber system. Some of these reasons are:

System scrubber chemistry has been very difficult to control; that is maintaining the limestone feed slurry fineness, concentration and feed rate within limits that do not produce scale. The reasons for this control problem will be explained later. When operated outside normal chemistry limits, the scrubbers generate large quantities of calcium sulfate scale that cause pluggage and erosion of venturi spray nozzles. Most of the scrubber modules have been operated with only the venturi recirculation circuits in service since start-up. The nozzle pluggage condition has led to poor fly-ash removal which has resulted in a rapid buildup of flyash deposits on moisture separators, reheaters, and ID fans. In addition to poor flyash removal, these periods of nozzle pluggage also permit some of the sulfur trioxide in the flue gas to pass through the scrubber, resulting in a condition where the gas entering the induced draft fans is below the SO_3 dew point, even after reheating the flue gas 25 F. This condition results in wet, acidic (pH 1) flyash deposits on ID fan and reheater surfaces. This combination of flyash and sulfuric acid has eroded fan blades and caused failure of the 304 stainless steel reheaters.

A test program is currently under way to determine the following:

1. Degree of flue gas reheating that will result in minimal deposits on the fans.
2. The best method of obtaining the additional reheat.

In the meantime, 75 degrees reheat has been temporarily provided by taking secondary air at 495°F from the air heater and blending this with flue gas leaving the scrubbers to obtain a 200 degree flue gas entering the fans. This reheat arrangement limits the unit load to approximately 650 MW.

"D" module is the only module that is complete with both venturi and absorber recirculation circuits, two stages of absorber trays and a water wash stage below the moisture separator. Testing during the past year indicates that dust loadings leaving this module are better by a factor of two than dust loadings obtained from the venturi-only modules. The average of 37 tests on "D" module (table 1) show that the dust removal performance has been better than predicted.

Within the last few weeks, one additional absorber stage and a water wash tray have been added to "C" module and tests were conducted to verify its performance. The results of 11 dust tests show better performance than either the venturi alone or the venturi and one absorber stage configuration. Additional absorber and water wash trays are being installed in the other modules as they become available for revision.

The sulfur dioxide removal performance at La Cygne is dependent upon many factors:

(1) module equipment (2) purity of limestone from the quarry and (3) operation of the milling system. This summary of SO_2 performance (table 2) show from 75-83% SO_2 removal has been obtained from "D" module when the above conditions are close to expected. Operation with one absorber stage has resulted in 50-74% SO_2 removal, and with the venturi only, we get 35-50% percent efficiency. (Table 2)

Proper chemistry control is the most important factor required for the successful operation of the

La Cygne scrubber system. This control not only includes maintaining recirculated slurry within certain operating limits but also encompasses the selection of raw limestone rock, reducing it to the right fineness, and feeding the limestone slurry to the scrubber at the right time.

The limestone and the coal at La Cygne are both mined locally; both vary considerably in chemical analysis; and if you don't obtain the right type, you are going to have problems. Unfortunately, with both the coal and limestone, you don't know you have the wrong type until it gets into the system. Figure 4 shows some recent variations in limestone purity since April of this year.

During the period April-August, 1974, we went through a period of excellent chemistry control, seldom having to take a module out of service due to low pH. This period the limestone purity was generally greater than 90 percent calcium carbonate, the magnesium carbonate concentration was less than 1.5 percent, and the milled feed slurry fineness in the range of 4.5-6 microns. A number of interesting observations were also made during this operating period:

1. The system pH was easily maintained at pH 5.7-6.0 with limestone feed at 150-175 percent of stoichiometric requirements.
2. Little scaling was observed in the scrubber modules.
3. The oxidation rate of calcium sulfite to sulfate was as low as 10-20 percent during some periods of operation.

During this operating period large hydrocyclones were put in service on all the module venturi recirculation streams. Prior to this time, it was almost impossible to prevent venturi pump recirculation screens from plugging with scale. These were replaced with low-pressure-drop hydrocyclones designed to remove particles greater than 50 mesh in size from the entire 5,500 GPM venturi recirculation flow. Oversize material is discharged into a grit screen located on top of the recirculation tanks. This system has reduced nozzle pluggage in the venturis and has greatly decreased erosion of valves and spray nozzles in the scrubber system.

With four to six ID fans operating in parallel from the plenum common to the discharge of all seven modules, some will assume all the load and others unload when the negative pressure reaches 42"-45"

H₂O. Unless lowered by the Control Operator this will lead to a boiler trip. Individual fan characteristics are currently being studied to correct this problem.

Present procedures call for cleaning one module each night on a rotational schedule and attempting to keep all seven in service during the day. Washing requires two to three men approximately ten hours to complete, including tagging and opening the many sections. The predemisters and absorber trays are normally clean unless scaled over from erratic pH control. The venturi area usually requires nozzle and wall wash cleaning which may include dismantling up to 7 or 8 of the 80 nozzles. The common sump under the venturi and absorber usually contains from 4 to 12 inches of hard sludge formed from flyash, carbonate, and sulphates and often is like concrete to remove. We currently are experimenting with this "puddle" to see if removing the sump screens will eliminate this fluid level.

COSTS

The operating expense for the first nine months of 1974 is as follows:

Operating labor	\$ 187,497	0.17 mils/kW
Operating materials	\$ 26,353	0.12 mils/kW
Maintenance labor	\$ 260,486	0.23 mils/kW
Maintenance material	\$ 456,567	0.42 mils/kW
Limestone	\$ 627,419	0.56 mils/kW
TOTAL	\$1,566,322	1.40 mils/kW

Capital Costs

Total Unit	\$195,000,000	\$238/kW
Total Scrubber	\$ 42,000,000	\$ 51/kW

The problems discussed were some of the more involved and time consuming that we confronted. Many repairs and some design changes have been made to the scrubber during the shutdown for first-year turbine inspection - February 15 to April 6. These changes have given us a much more reliable and maintenance-free operation. Startup problems are always expected, and we believe we now have the experience to maintain reliable operation, while contending with any other problems that may appear.

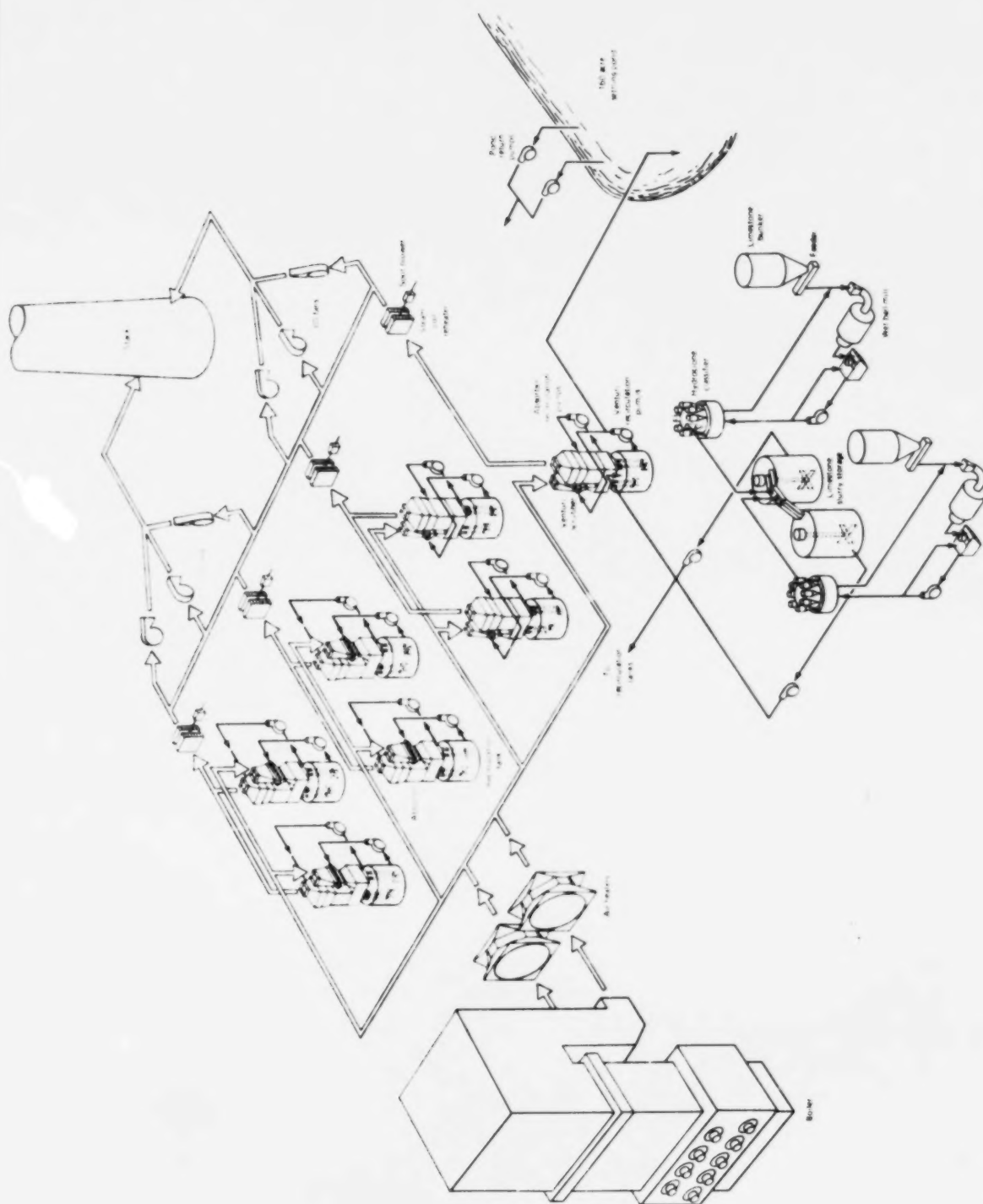
Improvements being considered for the system are:

1. Addition of fan eighth module
2. Additional forced draft fan
3. Additional induced draft fan
4. Separate conveyor system for limestone
5. Different ball mill liners - manganese or rubber instead of nihard.
6. Modernizing certain instrumentation for reliability. Add instrumentation to the moisture content of reheat.
7. With the remaining life of the 160 acre settling pond estimated at 1-1/2 years, evaluation will have to be made for an additional lake, removing the contents to fill in a nearby quarry or coal pits, or building the dams higher.
8. Module Improvements:
 - a. New improved demister installation
 - b. Additional absorber tray
 - c. New perforated predemister installation
 - d. Second demister tray installed in addition to new tray
 - e. Re-evaluation of reheat system to include reheat tubes and heat from an external source such as oil burners

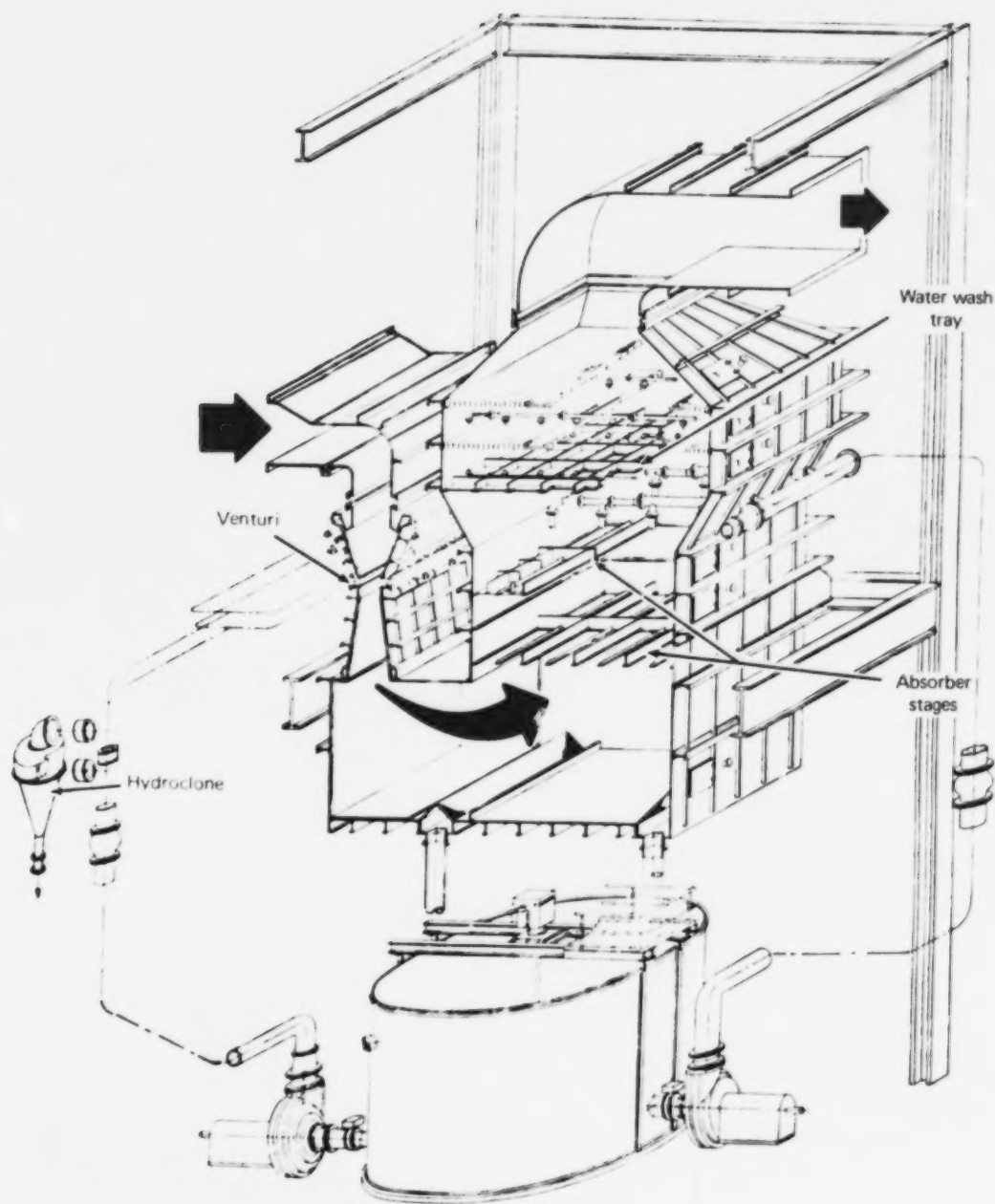
- f. Additional wash systems to eliminate periodic cleanings

Performance & Availability

The availability of the boiler-turbine has been increasing steadily, from 37% in January, to 60% in May, to 83% in September. This has consequently put increased pressure on continuous operation of the scrubbers. There are three demands affecting module outages: (1) regular cleaning (2) maintenance requiring immediate attention, and (3) continuing modifications. Often these cannot be done simultaneously but for the past few months module availability has been 100%; 7 modules have been operating 20%, 6 modules 90% and 5 modules 96% of the time the boiler is available. In all fairness to our past efficiency tests it makes a tremendous difference in the cleanliness of the module, the solid control through good instrumentation and maintaining the unit without imbalances.



La Cygne limestone wet scrubbing system



Single scrubber module

Particulate Performance

Module	No. of tests	Outlet dust loading lb/1000lb gas	Remarks
A, B, C	30	0.178	Venturi + 1 absorber stage
C	11	0.115	Venturi + 2 absorber stages + wash water tray
D	37	0.080	
		0.091	Predicted performance equivalent to EPA 0.1 lb/MKB

Table I

SO₂ Performance

Module	SO ₂ ppm		Removal efficiency %	Range of SO ₂ removal %	Remarks
	inlet	outlet			
A	4522	2710	40	35-40	Venturi
F	3970	1410	64	50-74	Venturi + 1 absorber stage
D	4453	819	82	75-83	Venturi + 2 absorber stages
				80	Predicted performance

Table II

Limestone analysis and fineness

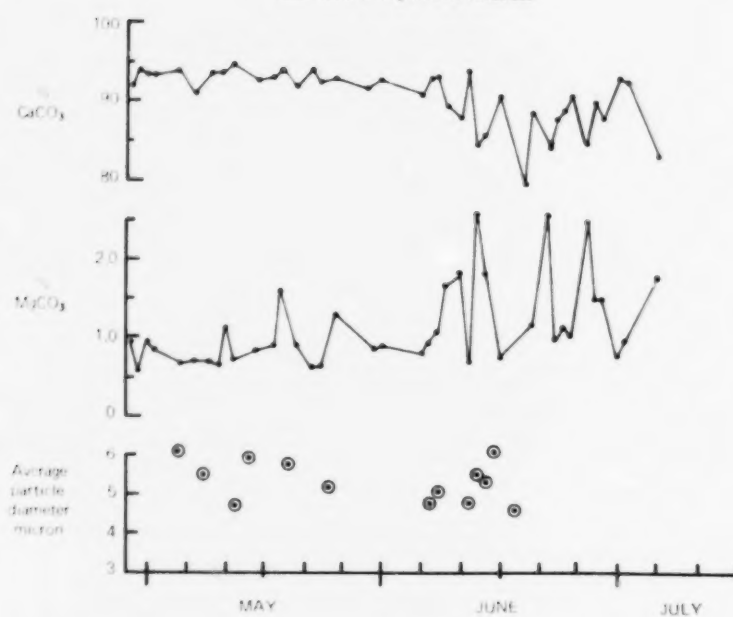


Figure 4

DUQUESNE LIGHT COMPANY
PHILLIPS POWER STATION
LIME SCRUBBING FACILITY

By

Steve L. Pernick, Jr
Manager-Environmental Affairs
Duquesne Light Company
Pittsburgh, Pennsylvania

and

R. Gordon Knight
Superintendent-Technical Services
Duquesne Light Company
Pittsburgh, Pennsylvania

Prepared for presentation
at
EPA Flue Gas Desulfurization Symposium
Atlanta, Georgia
November 4-7, 1974

Preceding page blank

329

DUQUESNE LIGHT COMPANY
Phillips Power Station
Lime Scrubbing Facility

CONTENTS

	Page
Introduction.....	331
Pilot Plant Operation	332
Installation Schedule.....	333
Description of System	335
Initial Operation	338
Extended Shutdown for Repairs.....	340
Second Start-up Operation.....	343
Sludge Treatment and Disposal.....	345
Future Operating Plans.....	348
Availability and Reliability.....	350
Capital and Operating Expenses.....	350
Future Objectives.....	351

DUQUESNE LIGHT COMPANY
Phillips Power Station
Lime Scrubbing Facility

INTRODUCTION

Duquesne Light Company is an investor-owned electric utility serving approximately 1/2 million customers in southwestern Pennsylvania including Pittsburgh. The Company has a net generating capability of approximately 2700 MW and presently operates three coal burning power stations, two of which, Phillips and Elrama, are discussed in this paper.

In 1969, Gibbs & Hill, Inc. was engaged to conduct a comprehensive study of the most feasible means of complying with: the emission regulations being promulgated at the time by Allegheny County, which has control over the Phillips Station; the regulations of the Commonwealth of Pennsylvania, which has jurisdiction over the Elrama Station, and the most stringent regulations that could reasonably be conceived. The study was completed in September, 1970. It represented a comprehensive evaluation of all test data concerning emissions from the two power stations and an intensive investigation of all possible means of compliance with the applicable and anticipated regulations. The study concluded that: (1) the use of low sulfur western coal was not a feasible means of compliance since it would not achieve compliance with the particulate emission limitations of the regulations with the existing dust collection equipment; (2) that conversion from coal to low sulfur oil was not feasible due to the uncertainties of obtaining a sufficient supply of the necessary large quantities of such oil; and (3) that the most practical means of sulfur dioxide emission control was desulfurization of stack gases. The consultants then conducted a thorough study of existing methods of sulfur dioxide removal from stack gases. At the time of their investigation in 1969 to 1970, only two full scale stack gas scrubbing systems were in operation in the United States. One of these has since been abandoned and the second has had a poor record of operating reliability. Furthermore, the injection of limestone directly into the boiler, which was the basis of these SO₂ removal processes, was not at all suitable for consideration at the Elrama and Phillips Power Stations since it would result in additional boiler tube erosion to which the boilers were already particularly susceptible. Furthermore, available data indicated that SO₂ removal efficiency might not be sufficiently high to comply with the regulations which were among the most stringent in the country.

Because of the characteristics of our boilers and our requirements for a system that would remove both particulates and sulfur dioxide, the report concluded that a tail-end scrubbing system was the best available means of control and recommended that we install a dual stage venturi scrubbing system. Some of the technical

reasons for choosing the system are as follows:

1. It was thought that a dual stage scrubbing system would enable separating the sludges resulting from scrubbing for fly ash in the first stage and scrubbing for SO_2 in the second stage, in the event the sludge from the SO_2 removal stage became a problem.
2. A dual stage scrubber had the potential of being converted to a magnesium oxide regenerative system in the event it became desirable or practical to do so.
3. Venturi scrubbing for particulate removal was fairly well proven in the steel industry. Furthermore, if only one of the scrubber trains was equipped with a dual stage SO_2 prototype, and this was proved to be impractical or unworkable, we would still have a workable single stage venturi scrubber system for particulate removal. We could then investigate an alternate method of SO_2 removal with minimal loss of investment.

In addition to the above reasons was the fact that preliminary time schedules indicated that these scrubbers could be delivered with the least delay. Duquesne accepted the recommendations of Gibbs & Hill, and in view of the unproven nature of the wet scrubbers for sulfur dioxide removal, decided that the most prudent approach to the installation was to begin by equipping only a portion of the scrubber system at the Phillips Station with sulfur dioxide removal equipment. Subsequently, Gibbs & Hill was instructed in September, 1970, to proceed with the designing and installation of the Phillips scrubber system. The scrubber installation at this plant was to include four scrubber trains. Three of these were single stage scrubbers for dust removal, and one was a dual stage scrubber (two single stages in series) which is the prototype for sulfur dioxide removal. Each scrubber was to be capable of approximately 540,000 CFM, which is the equivalent of approximately 125 megawatts. With the station rated at 387 megawatts, this meant that the station would be equipped with essentially a spare scrubber train. This was a criterion specified by Duquesne Light Company to assure sufficiently high availability of the system.

It was anticipated that after a trial period, the operation of the prototype would serve as a basis for decision as to the installation of additional similar SO_2 removal equipment at this station and other coal burning facilities to comply with the applicable regulations.

PILOT PLANT OPERATION

In February, 1971, a pilot plant size Chemico venturi type, dual stage scrubber system with a capacity of 1500 CFM was installed at the Phillips Power Station to study the performance and operating characteristics of the dual stage venturi scrubber system. The pilot

plant was connected to the duct entering the mechanical dust collector of No. 5 boiler. The scrubber was operated with various types of lime-stone and lime. In addition, other conditions such as the stoichiometry, liquid to gas flow ratio and pressure drop were varied in order to determine the conditions which would yield optimum collection efficiency and operating conditions. After determination of these parameters in approximately 3 months of operation, the pilot scrubber was dismantled on May 8, 1971.

In the meantime, in March, 1971, invitations to bid were issued by Duquesne Light Company to Chemico and Combustion Engineering for a scrubber system capable of removing sufficient SO_2 and particulates to comply with existing and anticipated future regulations. The bid was subsequently awarded to the Chemico Corporation in July, 1971. The battery limits for Chemico was limited to the scrubbers and associated pumps and controls between the inlet hot gas duct manifold to the exit wet gas header, including the reheater but excluding the new induced draft fans.

In obtaining approval of our plan, compliance programs were negotiated with Allegheny County for the Phillips Station and with the State for the Elrama Power Station. The program included the installation of five single stage venturi scrubbing trains for particulate removal only at the Elrama Station, to be installed concurrently with the Phillips scrubber system as described above. The plan was to use the prototype sulfur dioxide removal installation at Phillips as a basis for decisions concerning additional SO_2 removal equipment at both stations in order to comply with SO_2 emission limitations.

INSTALLATION SCHEDULE

Although in Duquesne's best judgment, the scrubber system could not be made operational, even for particulate removal, prior to July 1, 1973, which was 34 months from the date of our decision to install a scrubber system, Duquesne Light eventually acceded to pressure from the State of Pennsylvania and established a target date of January 1, 1973. Following startup, the schedule called for a two-month "debugging" period followed by a ten-month test program on the prototype SO_2 scrubber at Phillips. This would determine if stack gas scrubbing with alkaline injection on a plant scale basis was a feasible and reliable means of removing sulfur dioxide from the gases. The plan was to study operating conditions during this period to establish optimum SO_2 removal efficiency using various types of lime, the operating mode most conducive to reliable trouble-free operation, and other essential information. The schedule called for a decision to be made by January 1, 1974, as to whether or not the operation of the prototype SO_2 scrubber was satisfactory or if an alternate system should be investigated. If the operation of the prototype sulfur dioxide scrubber proved satisfactory, the schedule called for design, procurement and installation of additional SO_2 removal equipment by January 1, 1975, for both stations. At that time, compliance with County and State regulations would be achieved. (For a list of significant dates see Figure 5)

Despite our best efforts, no part of the Phillips scrubber installation was placed in operation until July 9, 1973, and then for only two boilers. At that time, two of the scrubber trains were placed in operation, and two of the six boilers were connected to the scrubber system. This was consistent with our startup schedule which called for connecting one boiler at a time to the scrubber system so as not to jeopardize the availability of a significant portion of the plant's generating capability in the event of failure of the scrubber system. The two boilers connected have a combined steam generating capability equivalent to approximately 80 megawatts.

The July 9, 1973 startup date represents a slippage of approximately 6 months from the original schedule. This slippage was attributable to the following factors:

1. The venturi scrubber system was not a completely engineered system at the time it was purchased, and a considerable amount of developmental engineering was required to adapt the scrubbers to a system that would meet our requirements. Chemico was also charged to incorporate, to the limits of their responsibility the latest available operating and maintenance experiences at the Four Corners Plant in New Mexico, the Dave Johnston Plant in Wyoming, Mitsui Plant in Japan, and any other Chemico installation having experiences adaptable to our proposed installations.
2. Equipment suppliers and fabricators did not meet promised delivery dates due to the volume of orders, decreased productivity, and inadequate quality control. Considerable money was spent to expedite and inspect equipment and materials for our project. In spite of this effort and expense, many late deliveries and mistakes in fabrication occurred. Some examples are as follows:
 - a. The induced draft fans were scheduled for delivery in Phillips by July 15, 1972. Delivery was not completed until December 12, 1972, a delay of five months. Not only were the fans late, but numerous cracks and defects in the welds on the fan rotors were found after delivery. It was necessary to grind out all of the defective welds and make the necessary repairs in the field.
 - b. An approximate five-month delay was experienced in delivery of the ductwork. In addition, because of the huge size and weight of the ductwork, many structural support schemes had to be studied before an acceptable method of reinforcing the existing station building to accept the unusual weight load of the ductwork could be determined. After a reinforcement plan was devised, errors were discovered in the implementation of the plan, requiring corrective design work.

- c. Delays of up to six months were experienced in the delivery of electrical equipment such as switchgear and transformers. In addition to the delay of equipment, errors were found in the wiring diagrams for this equipment. Once the equipment was received, design errors were also discovered, such as improperly designed bus connections on the switchgear and improper alignment of bus bars on power centers. In addition, failures of switchgear and transformers were experienced during field testing.

These are only a few examples of the numerous delays and errors by equipment suppliers which affected the final completion date of the first phase of the Phillips project.

- 3. Major delays were also incurred in the field, such as the following:
 - a. Because of engineering delays and late delivery of equipment, construction schedules had to be continually revised which seriously affected productivity.
 - b. Space limitations existing at the job site, more serious than anticipated, affected productivity because of the lack of accessibility and maneuverability. Special equipment and construction methods for the installation of the equipment were required beyond those originally anticipated, thus contributing to additional delays.
 - c. The availability of boiler outages for tie-in of the new ducts, presented additional problems. Outages had to be scheduled months in advance, and any revision in schedule was dependent upon availability of purchased power and the outage schedules of other generating equipment, both on our system and on interconnections.

DESCRIPTION OF SYSTEM

Design of both the Phillips and Elrama systems was a joint effort by Gibbs & Hill and Chemico, each having certain battery limits. Construction was the responsibility of Gibbs & Hill. Only the Phillips system will be described in detail.

The Phillips system comprises four parallel trains, Figure 1, and was designed to handle a total gas volume of 2,190,000 ACFM with all trains in service. With one train out of service, the three trains have a capacity of 1,600,000 ACFM, which is sufficient for 400 MW, assuming normal excess air.

Three scrubber trains were provided with first stage scrubbing for fly ash removal, Figure 2. The fourth train, known as the "prototype" SO₂ train, was fitted with a first stage particulate removal scrubber and a second stage SO₂ absorption scrubber as shown on Figure 3.

The ductwork on each boiler was modified to enable operation with the gases going either to the scrubbers or to the original gas path after passing through the existing dust collection equipment. A mechanical and electrical dust collector are located in series ahead of each individual boiler stack. The existing ducts leading to the old ID fans and stacks were blanked off with removable steel plates to enable diversion of the flue gas to a new common duct. The new ductwork is routed to the scrubber building where 16 foot diameter takeoffs lead to each of the four first stage scrubbers.

The hot flue gas (about 340°F) containing fly ash and sulfur dioxide (about 1400 ppm SO₂) enters the first stage scrubber and impinges upon the upper cone. One half of the total scrubber liquor (8250 gpm) is introduced, Figure 4, into the vessel through the bull nozzle where it is sprayed over the surface of the upper cone, to initiate the scrubber action. A second stream of 8250 gpm scrubber recycle liquor enters through the tangential nozzles at a point above the adjustable throat damper. The flue gas and scrubbing liquor are brought into intimate contact in the throat section of the scrubber where the particulates and some SO₂ are removed. The gas and liquor continue downward to the separator section where, after being separated from the scrubbing liquor, the flue gas enters a demister where entrained liquor is removed. The gas leaves the scrubber and enters a new wet ID fan which is provided with water sprays to remove any accumulation of solids resulting from carryover from the scrubber. Fresh water is used for spraying.

The ID fan housings are lined with 1/4 inch thick natural rubber. Wheel material is Carpenter 20 Cb 3. The shaft is 316 L stainless steel. Each fan is driven by a 5500 HP, 1200 RPM, 4160V electric motor. Bearings are of the water-cooled type, served from a closed cooling water system.

Outlet gas temperature, normally 110-120°F, from the first stage scrubber is monitored. At 175°F a control valve is automatically opened to admit emergency cooling water to the upper cone. Additional temperature rise would automatically shut down the fan and close the isolation dampers.

The gas leaving the ID fan enters one of two vessels depending on the train. In the case of the three single stage scrubber trains, the gas enters an entrainment separator in which entrained water is separated and collected. The separators or mist eliminators are separate vessels which will be replaced by second stage scrubber vessels if the test program indicates this to be necessary. In the

case of the SO₂ prototype train, gas leaving the ID fan enters a second stage scrubber vessel, which is identical in size and design to the first stage. This second stage vessel is equipped with a reagent (lime) injection nozzle in the bottom cone. The scrubbing liquor is then picked up by the two recycle pumps (Carpenter 20 Cb 3). Some of the liquor is bled off to the first stage scrubber, where some is removed from the cycle to thickeners, and the balance is recycled.

Total recycle flow is about 16,500 gpm with a 615 gpm bleed to the thickeners. All single stage vessels have also been fitted with reagent nozzles and feeds to enable control of pH and incidental SO₂ removal.

The exit gases leaving either the second stage scrubber vessel or the mist eliminator vessel enter a common wet duct lined with Flakeline 103 (a product of the Ceilcote Company consisting of a glass flake filled modified polyester resin, subsequently referred to as Ceilcote) which leads to a new concrete acid resistant brick lined chimney. Prior to entering the 340 foot chimney, a section of the wet gas duct, constructed of 316L SS is equipped with a reheater, employing a direct oil-fired burner system. The two-burner system is capable of reheating 30°F, although the intent is to normally reheat 20°F.

Bleed off from all scrubbers is directed to a trough feeding two 75 foot diameter thickeners. The overflow from these thickeners is directed to a collecting tank where it is pumped by two-2000 gpm SS pumps to the make-up line for return to the scrubber system.

Thickener underflow, at 35 to 40% solids concentration, is pumped by one of two 105 gpm, 15 HP pumps to one of three clay lined sludge holding ponds. Each pond has a capacity of about 6500 cubic yards. Just prior to discharge to the holding ponds, the sludge enters a mixing tank where a stabilizing agent can be added at a predetermined rate based on the density of the sludge as it leaves the thickener. After the additive is intimately mixed with the sludge, the mixture discharges by gravity to one of the ponds.

In the pond, both settling and curing take place. Supernatant liquid is withdrawn and recirculated to the scrubber system via the thickener. The sludge curing ponds provide interim storage while stabilization is taking place. One pond received thickener underflow, while a second is curing and the third is being excavated for final off-site disposal.

Lime is fed from a storage silo at a controlled rate to the lime slaker where it is slaked with fresh make-up water. The slaked lime overflows into a slaker transfer tank, where make-up water is added to provide a constant flow of lime slurry with a concentration of about 15%.

A service (river) water system independent from the power station service water system is provided. It includes a pair of 900 gpm, 100 HP pumps taking suction from the existing condenser discharge tunnel and a 10 inch distribution header for the scrubber facility. It provides fan spray water, pump seal water, mist eliminator and scrubber demister spray water, instrumentation flush water, chemical mixing tank, and emergency water for scrubbers.

INITIAL OPERATION

Although the lime addition system was not yet completed, partial start-up of the scrubber system began on July 9, 1973. The decision to proceed with startup, without the lime addition system, was made in light of the fact that we thought the scrubber system was suitable for operation at a low pH. Almost all of the major components and piping were either Ceilcoted, rubber lined or fabricated of a corrosion resistant material to withstand a low pH. During the first five days of operation, a pH of approximately 1.3 to 2.0 developed in the scrubbing liquor while burning 1.8-2.0% sulfur coal. Because of pressure to proceed with the debugging operation, and because of our reluctance to incur additional delays, the decision was made to continue operation with the low pH rather than shut down the scrubbers until the lime system became operational. Unfortunately, as a result, the scrubber system was forced out of service due to the corrosion failure after five days of the two thickener overflow return pumps, which were intended for operation at a pH of 5 to 6. They were promptly replaced with two 316 L SS pumps. Use of the lime system feed to the thickeners was obtained after the first week of operation.

During the first two weeks of operation, additional problems developed and were corrected in a two-week outage following the first two weeks of operation:

1. A piece of rubber liner was found in the suction of one of the recycle pumps. The individual scrubber was shut down in an effort to locate the source; however, it could not be determined until the corrosive nature of the scrubber media caused the failure of a 20"/16" Tee within 10 days.
2. Leaks occurred in the lead floor lining of the stack which were a potential threat to the stack concrete structure. The lead sheets had been soldered rather than "burned." The lining was replaced.
3. Leaks developed in four 405 SS expansion joints in the gas ducts leading to the stack. The appearance of the failed metal was variously described as that of "lace curtain" or "swiss cheese." This resulted from partial-load type of operation of the scrubber system requiring us to partially open the vacuum breakers and allow ambient air

to supplement the gas flow through the scrubbers. Condensation resulted. The joints were replaced with an asbestos/butyl material.

4. The reheater, which utilizes direct firing of No. 2 oil in the cold gas duct as it enters the stack, is not yet operational. We have been unable to obtain satisfactory operation of the reheater due to burner problems and oil pump problems. We have placed no priority on solving these problems due to our reluctance to operate the reheater considering the present energy crisis and oil shortage. Whenever the reheater becomes operational, we plan to operate it during periods of adverse meteorological conditions to obtain a higher effective stack height.
5. During the outage, evidence of corrosion was found on the ID fan shaft shrouds, inlet dampers and stiffener bars - all 316 L stainless steel. Two corrective measures were taken - redesign of the fan sprays and application of 1/4-inch rubber coating to the affected parts. After two weeks of operation the chloride concentration in the liquor (closed cycle) had reached 800-900 ppm. This, no doubt, contributed to the corrosion noted above. It was eventually to reach 3500-4500 ppm.

On August 5, 1973, the scrubber system was returned to service with the lime feed system for pH control continuing in operation. In addition to No. 1 and No. 2 boilers, the No. 3 boiler was also connected to the scrubber system, which made a total of approximately 145 megawatts.

During the following weeks of operation, we were informed by Chemico of a major change in the operating requirements of the scrubbers; namely, the necessity to bleed more fluid to the thickener tanks so that the concentration of solids in the recycling fluid could be kept to a minimum in order to protect the recycle pumps, and to minimize the settling of solids in the bleed lines. This caused a change in the design concept by essentially doubling the bleed rate. This meant that the two 900 gpm thickener tank overflow return pumps, which had been provided with the system, were inadequate since they no longer had sufficient capacity to supply the needs of the scrubber system under one pump operation (the second being a spare) with all boilers connected to the scrubber system. Two 2000 gpm pumps were obtained for later installation in parallel with the two 900 gpm pumps.

During the two months of operation following the August 5, 1973 start-up date, operation of the scrubbers was alternated between trains. In other words, as problems arose on one train, it was shut down and another train was put in service to take its place. No effort was made to operate the dual stage prototype SO_2 scrubber for full SO_2 removal because it was felt that debugging operations could more effectively proceed with the most basic operation possible, in other words,

with the single stage scrubbers.

EXTENDED SHUTDOWN FOR REPAIRS

During the operation of the scrubbers in August and September, 1973, additional problems came to light which caused an extended shutdown of the scrubber system, and a further delay in the operating schedule, namely:

1. A routine inspection of the new stack revealed additional moisture leaks from condensation (pH 2.0) seeping through the mortar joints and running down the outside of the acid brick lining. Seepage was occurring in more than 100 areas of the brick lining. The condensate was collecting at the bottom of the stack in the annulus between the lining and the shell. The acid condensate was also seeping through construction joints and down the outside of the concrete shell permitting it to attack the concrete, which could possibly impair the structural stability of the stack. The contractor advised us that a four week outage would be necessary to permit inspection of all mortar joints and to repoint as required. This outage for other reasons to be described, extended for about five months.
2. Significant scaling and buildup had occurred in various portions of one or more of the scrubbers and in the associated piping in varying degrees as described below:
 - a. Buildup was found on many of the upper cone support struts. These football-size accumulations appeared to be creating a disturbance in the water flow pattern over the cone resulting in accumulations of deposit on the cone surface.
 - b. Deposits were found in the throat area of two of the scrubbers, sufficiently thick in some areas to prevent the maintenance of the six inch design pressure drop across the throat.
 - c. A thin buildup of hard scale was observed on the face of the throat dampers of one of the scrubbers.
 - d. Deposits of fly ash were found in and around the operating mechanisms of the adjustable throat dampers. These were sufficient to prevent full adjustment of the dampers.
3. Corrosion/erosion was found in numerous areas:
 - a. Severe pitting and corrosion appeared on all of the 316 L clad ID fan damper blades and frames. The cladding was

removed and the carbon steel was Ceilcoted.

- b. In all scrubbers, a moist layer of fly ash was observed on the carbon steel skirt that directs the flue gas down the cone. Extensive rusting was found under the fly ash and on the underside of the vessel tap as well. The latter may have resulted from frequent outages for inspection, during which alternate wetting and drying occurred.
 - c. The upper cones of the scrubber vessels were fabricated with a 316 L SS apex. The balance was carbon steel with a Ceilcote coating. The uncoated apex showed erosion/corrosion in all vessels. Similar attack was shown on the inner surface of the 316 L bull nozzle directly above the cones. Both affected areas were coated during the outage.
 - d. Baffle plates in some tangential nozzles (12 per vessel) were found badly corroded. The specified material of construction was 316 L SS, but it was found that the corroded baffles had been fabricated of 304 SS. Sixteen baffles out of a total of 60 required replacement with 316 L.
 - e. Similarly, the throat damper arms (12 per vessel) were supposed to be 316 L SS. Thirty-two out of a total of 60 were badly corroded and were found to be 304 SS. They were replaced with 316 L.
 - f. The top and bottom throat damper scraper blades were practically destroyed. They also were found to have been 304 SS instead of 316 L. They were replaced with 316 L.
4. An initial inspection and analysis of the ID fans by Franklin Institute Research Laboratory (FIRL) indicated the existence of chloride stress corrosion cracking of the parent metal and structural welds, as well as general corrosion on all exposed surfaces. At that time, they advised us that the problems resulted from the hostile environment in which the fans were operating. In addition, they made a specific recommendation that the fans not be operated in the condition found at that time, due to possible catastrophic failure.

As a result of their search for protective coatings, FIRL recommended the application of a Ceilcote material, Coroline 505AB. This is a hard, impervious, acid resistant, epoxy base material which is applied at a nominal thickness of approximately 1/4". The application of the Coroline is a tedious, time consuming hand application by trowel and requires close

quality control. The application of the material began in November, 1973. Although Coroline had not been extensively used on fan blades, or proven to any degree, we were given reasonable assurance that it would be successful. As a result, we began application of this material to as many fans as manpower permitted.

On January 22, 1974, the first fan was completely coated and was started up for balancing. After a balance move was attempted, the fan was forced out of service due to excessive vibration resulting from separation of a portion of the coating from one of the blades. The Ceilcote Company informed us that the separation was localized and probably caused by a faulty batch of material which did not properly adhere to the metal. The Coroline was re-applied and the fan was started again on February 5, 1974 and again was forced to shut down due to separation of the material in a second area. At that time, the Ceilcote Company was unable to adequately explain the failure of the coating to adhere to the fan surfaces. As a result, the decision was made that the application of Coroline as protection against the adverse effects of the flue gases was not the answer to the problem. It is believed that the flexing of the blades during operation may have contributed to the lack of adhesion between the Ceilcote and the blades.

By this time, FIRL had been able to conduct more exhaustive investigations and analyses on the fan blade material. As a result, they revised their initial conclusion and reported that the fans were suffering from a much lesser degree of stress corrosion cracking of the weld metal and a degree of pitting attack on all fan blades. The report also indicated that fan failure was no longer considered imminent.

In review of the ID fan problems, it appears that the corrosion problems which we had experienced might be partially due to the operation of the scrubber system at a lower pH and a higher chloride concentration (3500-4500 ppm) than had been foreseen by the scrubber designers. Several steps were taken to try to alleviate the situation:

1. Operation of the dual stage scrubber with full lime addition, which we felt would be adequate for maintaining optimum pH control in that train;
2. Addition of supplemental lime feed to the other operating scrubbers, and
3. The installation of a redesigned spray system, which had proven effective at other installations.

These steps in addition to frequent scheduled fan inspections were felt to be adequate safeguards for placing the system back in operation.

Our ability to start up the scrubber system and initiate full SO₂ removal in the dual stage prototype was further improved by the completion of the sludge stabilizer addition system. We had been reluctant to add large quantities of lime to the single stage scrubbers, and equally reluctant to begin operation of the dual stage scrubber, both of which result in the production of considerable amounts of sulfate sludge, without the capability of adding a stabilizer.

SECOND START-UP OPERATION

Removal of the Coroline material from the dual stage scrubber fan No. 1 was completed, and startup of the scrubber system once again began March 17, 1974. At that time, the dual stage Scrubber Train No. 1, and Scrubber Train No. 4 were placed in operation with boilers No. 3 and 4 connected to the scrubber system. Since then there have been these additional developments:

1. Excessive Fly Ash

The sludge handling system was designed so that after 30 days of normal operation with all boilers tied in and with the dual stage SO₂ prototype and two of the three single stage scrubbers, the first curing pond should be approximately full. However, after the first two weeks, one pond was completely filled with only two boilers connected to the scrubber system during the first week and three boilers connected to the system during the second week. This means that, with an average of 120 megawatts connected to the scrubber system, we were producing approximately 7,000 tons of sludge in a two week period. Or, stated another way, if the entire plant were connected to the scrubber system, we would fill a pond within a week. The major factor contributing to this change in capacity proved to be an increased fly ash loading of the scrubber system. The sludge disposal system was based on a 90% combined collection efficiency of the mechanical dust collectors and electrostatic precipitators. The transition of the new scrubber ductwork to the old ductwork is such that the gas flow pattern of the gases through the precipitators proved to be greatly disrupted. Velocity traverse data indicated that laminar gas flow no longer exists. Rough estimates indicated that the amount of particulates leaving the precipitators and entering the scrubber system may have trippled. Additional details on the sludge handling will be covered later.

2. ID Fan Stress

As part of the continuing investigation to determine the conditions of the ID fans, Structural Dynamics Research Corporation (SDRC) was engaged to conduct a series of strain gage tests to determine the structural stability of the fans. Their tests indicated that the yield strength was being exceeded in several portions of the fan blades, and that a degree of metal deformation was taking place. After additional strain gage testing, SDRC recommended the installation of doubler (reinforcing) plates on each of the blades to reduce the stresses to acceptable levels. After actual tests with stressed welded specimens in the fan atmosphere, it was also recommended that the doubler plates be welded with an Inconel 112 rod, rather than the Carpenter 20, 4 N1A or 8 N12 rods previously used. Since the installation is on a fan-by-fan basis, so that a spare scrubber train would always be available for emergency backup, it is expected that all four fans will not be completed until the end of November, 1974.

3. ID Fan Attack

Frequent shut down of the ID fans for inspection revealed that significant attack was still taking place. Despite the addition of lime for SO₂ removal and pH control, a low pH was being encountered in the ID fan drains. This results from additional SO₂ scrubbing occurring with the fan spray system. In a trial installation, caustic was added to the fan spray water. However, an inordinate quantity is required to obtain even a pH of 4, and the trial will soon be ended. As another attempt to alleviate the attack, a new type of fog nozzle was installed on one of the ID fans in an attempt to increase the effectiveness of spraying. No evaluation is yet possible.

4. ID Fan Deposits

In addition to the cleaning being accomplished by both the original and the new fan spray systems, it was necessary to periodically remove the ID fans from service to remove deposits by manual methods. In recent weeks, this phenomenon has not required the need to clean the blades as frequently.

5. Recycle Pumps

An unsatisfactory degree of wear (about 30% in three months) is being experienced on the recycle pumps to the point where pump parts are being diverted from the Elrama scrubber system

to Phillips at such a rapid rate that the stock of installed pumps at Elrama will soon be depleted. Experimental pump impellers and wear rings of three different materials and design will be installed.

6. Thickener Capacity

Modifications were made to the thickener tank overflows to increase their hydraulic capacity and thus compensate for the increased dust loading and provide spare capacity.

7. Throat Dampers

Significant unacceptable accumulations are still being found in and around the scrubber throat damper operating mechanisms. These have restricted the damper operation and resulted in operation at pressure drops of up to 10 inches instead of the normal 6 inch drop. Cleaning requires removal of the train from service for manual removal of the mechanism and hand cleaning. About 288 manhours per scrubber are required.

8. Sump Pumps

The ID fan spray sump pumps have been wearing out much faster than anticipated. This is due in part to the low pH and to the higher solids concentration in the fan drains. Once again, we have had to draw on the Elrama installation for replacements.

9. Closed Loop Operation

A closed-loop system may be possible at maximum load with all boilers cut-in. However, the Phillips Power Station is not a base load station, and as a result, the load fluctuates between 30% and full load. At the lower loads, less water is evaporated from the system, while many of the points of water addition (purge, seal and spray water) continue. This results in an excess of water in the system. Temporary permission was obtained from the Pennsylvania Department of Environmental Resources (PDER) to discharge the excess to our existing bottom-ash settling ponds. This permits the settling out of particulate matter and dilution of the blowdown prior to discharge to the river.

10. SO₂ Guarantee Performance Tests

In July, 1974, several tests were made on the dual stage prototype scrubber for SO₂ removal, with inlet SO₂ concentrations ranging from 1263 ppm vol. to 1,603 ppm. at stoichiometric ratios from 1.01 to 1.15 the SO₂ removal ranged from 86% to 93%. The guaranteed SO₂ removal is 80%.

SLUDGE TREATMENT & DISPOSAL

At the time the present period of operation started, the sludge additive system had been completed to the point where it was considered operable. However, it soon developed that the pumping system for the additive was operating below design capacity. The

necessitated pumping a slurry with higher solids content, and the 750 ft. long transfer lines were frequently out of service due to plugging.

This stabilization additive chosen for use is Calcilox, a proprietary product of the Dravo Corporation. This material was developed in the approximate period of October through December, 1972, when we re-installed the Chemico pilot scrubber for the sole purpose of making sludge. Industry knowledge of sulfate-containing scrubber sludge and means of stabilization appeared almost totally lacking at that time. The scrubber was connected so it could be supplied from either upstream or downstream of the existing dust collectors of the No. 4 boiler to enable experiments with both clean and dirty gas. The material developed, Calcilox, is of the consistency of dry cement and, on the basis of laboratory tests, has the property of consolidating the sludge so that it can be disposed of as land fill without an objectionable leachate. The degree and time of consolidation depends on the amount added (3-15% by dry weight of sludge), the pH of the sludge (10.5 appears suitable for our curing time), and temperature (lower temperatures slow the process). Since actual field leaching tests had not been made, we installed two membrane-lined ponds in an inactive ash disposal area to enable collection and retention of leachate from Calcilox-treated sludge. Bottom drain systems were provided with valves to enable leachate sampling at intervals. One of the ponds has 22,860 yd³ capacity, the other 5720 yd³. These capacities may be increased by 50% if the material proves suitable for successive layering and compacting.

During the previously noted difficulties with the additive system, Calcilox was either added sporadically or not added at all. As a result, the sludge after residence time in the ponds was of a "soupy" nature, which made it difficult to excavate with a clam shell and to transport in open trucks. Prior to placing the material in the trucks, fly ash from the ash silo was deposited in the rear of the truck to seal the tailgate. Because of the difficulties in handling the sludge, additional boilers could not be connected to the scrubber system since it would tax our ability to remove the sludge from the ponds as rapidly as it was being produced. Since the sludge did not have the amount of additive which would normally be added prior to discharging it to the ponds, no appreciable degree of stabilization took place in the ponds. We did not feel that the "soupy" material was representative of that which would result from the normal sludge disposal system, and were reluctant to use one of the lined test ponds for this material. Therefore, we requested interim permission from the State's Solid Waste Disposal Division of the PDER to dispose of the material on the ash disposal area where we normally dispose of the fly ash and bottom

from the generating plant. The State consented to our request, and we started disposal of the unstabilized sludge on our ash disposal site. It was spread in thin layers, mixed with additional fly ash and then compacted.

Handling of the sludge was extremely difficult. The trucks could only be loaded to approximately 1/2 capacity because of the danger of the sludge flowing over the sides if the truck stopped or started too abruptly or made turns at excessive speed. In addition, because it was impossible to prevent sludge from dripping onto the sides of the truck during the truck loading operations, it was necessary that the trucks be washed prior to leaving the plant premises. Even with such precautions, occasions arose when sludge was deposited on the roadways. After several warnings from the local police and the Pennsylvania Department of Transportation, it was necessary to rent a street cleaner and a water sprinkler truck to clean the streets during the hours of sludge hauling.

In approximately May, 1974, modifications were completed on the additive system, and Calci'ox was then added on a continuous basis to the sludge with around-the-clock technical supervision. Filling time per pond was about 10-14 days with the increased quantity of fly ash, instead of the 28-30 days which had been planned. Similarly, the curing time was reduced from 30-60 days to 14. However, with about 10% Calcilox addition the sludge did consolidate to the extent that drag-line excavation was possible, truck tailgates did not require sealing, and it was possible to fully load the trucks. Observations of the sludge during excavation showed that although the top layer resembled hard clay in consistency, it became earth-like and then almost fluid near the bottom. It was necessary to mix the lower-most layer with dryer material from the upper layers to obtain a satisfactory handling quality.

The consolidated material was trucked to and deposited in one of the lined areas for leachate monitoring. However, the thixotropic nature of the material was such as to prevent either leveling or successive layering of truckloads. It was necessary to tailgate the sludge and then let it stand (cure?) for about six more weeks before it could be leveled, and dozers and trucks did not get "hung-up." This, of course, slows the disposal process considerably.

Our present procedure is to add consolidated sludge to the lined area whenever possible. At other times, it is mixed and compacted with dry fly ash on the normal disposal area. We have now been monitoring the bottom drains from the lined area at weekly intervals for six weeks. In that time, the total dissolved solids (TDS) have varied from an original

2203 mg/l to the present (October 2) 3380 mg/l. A sample analyzed by the PDER on September 22 verified the results on our own sample of that same date. No record of total volume of leachate has been maintained, and it is realized that this quantity can affect the measured TDS appreciably. Similarly, it is realized that the method of disposal by tailgating and our inability to level the top of the sludge to a possibly impervious surface has also had an effect.

In addition to the above tests, leachate studies were conducted on a more closely controlled basis. About June 1, 1974, we used a concrete mixture to prepare four sludge mixtures with various additives. The additives were fly ash, 2% hydrated lime, 5% hydrated lime and 11% Calcilox. Each mix was then added to a plastic wading pool which had previously been fitted with a valved underdrain system for leachate sampling. Since the tests started, we have added water to each pool once each week so as to cover the sludge after first sampling and draining all water from the previous week. After fourteen weeks, the most meaningful tests are those with the 5% lime and the Calcilox. They show that the total dissolved solids (TDS) on the lime test started at 1450 mg/l and are now 460 mg/l. The TDS on the Calcilox started at 3370 and are now 2000. Penetrometer readings in that time have increased to 2.7 tons/ft² on the lime and 4.5+ on the Calcilox. In summary, while more salts are leached from the Calcilox mix, the degree of consolidation is higher than that of the lime.

FUTURE OPERATING PLANS

Due to improvements in the method of sludge treatment and handling, and due to close control of the thickener tank underflow operation, it was possible to cut-in a fourth boiler to the scrubber system on August 18, 1974. This increased the capacity connected to the scrubber system to approximately 180 megawatts. Our plan is to operate the scrubber system with the four boilers for a period of time to determine the extent of any aggravation of existing problems and the occurrence of any new problems. If the problems become intolerable, it may be necessary to return the scrubber system to a three boiler operation. However, if the problems are not severe, we anticipate the possibility of adding a 5th boiler to the scrubber system. We are extremely cautious in our optimism because of the number of unknowns in the sludge disposal system such as the following:

1. Since the lime feed system is essentially operating at maximum capacity, additional boilers will reduce the pH in the scrubber system. We are told that a reduced pH would adversely effect the stabilization of the sludge. However, we do not know to what degree it will be affected. Also, if additional boilers are connected to the scrubber system,

the rate of sludge production may not permit optimum stabilization period which now appear to be greatly in excess of 14 days.

2. With the oncoming cold winter months, it is not known to what extent low temperatures will hinder stabilization.
3. If additional boilers are connected to the scrubber system, and it becomes impossible to optimize stabilization of the sludge prior to removing it from the ponds, it may be necessary to again seal the tailgates with fly ash prior to loading the trucks with sludge. This would then be conducive to sludge material dropping on to the road, freezing and causing complaints from the local citizens.

If we can connect five out of the six boilers to the scrubber system by November 30, 1974, we will begin our ten-month test program at that time to evaluate the efficiency, reliability, and practicability of the scrubber system. A decision would then be made on September 30, 1975, concerning the installation of additional SO₂ removal equipment for compliance with the SO₂ emission limitations. The number of additional second stage scrubbers required for compliance with the regulations will be determined during the ten-month test program, and this will establish the completion date for installation of all additional scrubbers.

In the hope of obtaining higher efficiencies of SO₂ removal with the use of lime, Duquesne Light is conducting tests using Thiosorbic lime (a Dravo development) which is a lime having a concentration of magnesium oxide in the order of 4 to 6%. Various tests on pilot scrubber systems indicate that the higher reactivity of such a lime results in SO₂ single-stage removal efficiencies in the 90's. In preparation for these tests, one of the single stage scrubbers was isolated along with one of the thickener tanks. This isolation was necessary to establish and maintain the desired chemistry, since lime slaking capacity is insufficient to maintain the necessary conditions in all scrubbers and both thickeners. A test run with Thiosorbic lime was made in September, 1974. However, we encountered difficulties in achieving a sufficiently high rate of utilization of the magnesium oxide. A few cursory tests under these conditions indicated SO₂ removal efficiencies somewhat better than those with high calcium lime, but also significantly less than anticipated. The tests were suspended later in September to further evaluate means of increasing the reactivity and utilization of the Thiosorbic lime. Resumption is planned for about October 21.

As of August, 1974, construction on the Elrama scrubber system, which is essentially identical to the Phillips scrubber system, had progressed to a point where no additional meaningful work could be accomplished until the problems with the Phillips scrubbers are resolved. With the modifications and test programs now being conducted at Phillips, it is anticipated that solutions to the present problems may progress to the point where startup of the Elrama scrubber system may be possible in May, 1975.

AVAILABILITY AND RELIABILITY

The availability of the Phillips scrubber system is meaningless at this point since we have essentially 100% backup. In other words, there are two spare scrubber trains which can be placed in service whenever a problem arises on an operating train. The service hours for each train for the period from March 17, 1974 through September 30, 1974 are 3430 hours for No. 1 train (dual stage), 1678 hours for No. 2 train, 1931 hours for No. 3 train, and 2803 hours for No. 4 train.

Reliability was built into the design of the system which called for a spare scrubber train enabling us to carry almost full station load on three scrubber trains. In addition, a degree of reliability is built into each scrubber train through the installation of spare pumps, thickener capacity and backup electrical controls.

We do not view bypassing as the solution to maintaining acceptable reliability. However, we have incorporated a manual type bypass into our scrubbing system as the best approach to avoid the serious consequences of scrubber failure. The system uses blanking plates to redirect the gas flow to the original stacks, and the boilers have to be shut down during the plate change. Under normal circumstances, this process would consume several days depending on the availability of manpower. There really was no alternative in designing a bypass system for our plants. Mechanical bypass systems utilizing a damper arrangement require tight dampers, and sufficiently tight dampers have yet to be developed. We also had serious space limitations on installation of a mechanical damper system.

CAPITAL AND OPERATING EXPENSES

As of this date, the scrubber systems at Phillips and Elrama represent a capital investment of approximately \$61 million. It is expected that an additional \$19 million will be required for equipment necessary to comply with the SO₂ emission limitations. This will represent a total investment of \$80 million or approximately \$91 per kilowatt, exclusive of the cost of any additional sludge disposal facilities required.

A recent estimate indicates that the annual operating and maintenance expenses, including fixed charges, at Phillips and Elrama with full SO₂ scrubbing will be in excess of \$30 million per year, or 5.5 mills per kilowatt hour, exclusive of additional sludge disposal facilities which will be required for full SO₂ removal operation. These expenses would represent an increase of approximately \$11 per ton of coal consumed. Expressed in different terms, these expenses, without fixed charges, would represent approximately a 50% increase over the actual operating and maintenance expenses for 1973.

In an effort to estimate the total capital and operating costs that would ultimately be incurred for the scrubber systems at the Phillips and Elrama Stations, including provision of adequate sludge disposal facilities for full SO₂ removal, a preliminary evaluation was made of the additional property required for development into a disposal area and the additional expenses involved in disposing of the sludge. This estimate indicates possible capital expenditures of \$110 million for scrubber installation and property acquisition and development, of which \$61 million has thus far been spent. This represents an investment of approximately \$124 per kw.

Present estimates of the sludge disposal costs are \$7 to \$10 per wet ton, or \$15 to \$20 per dry ton of sludge.

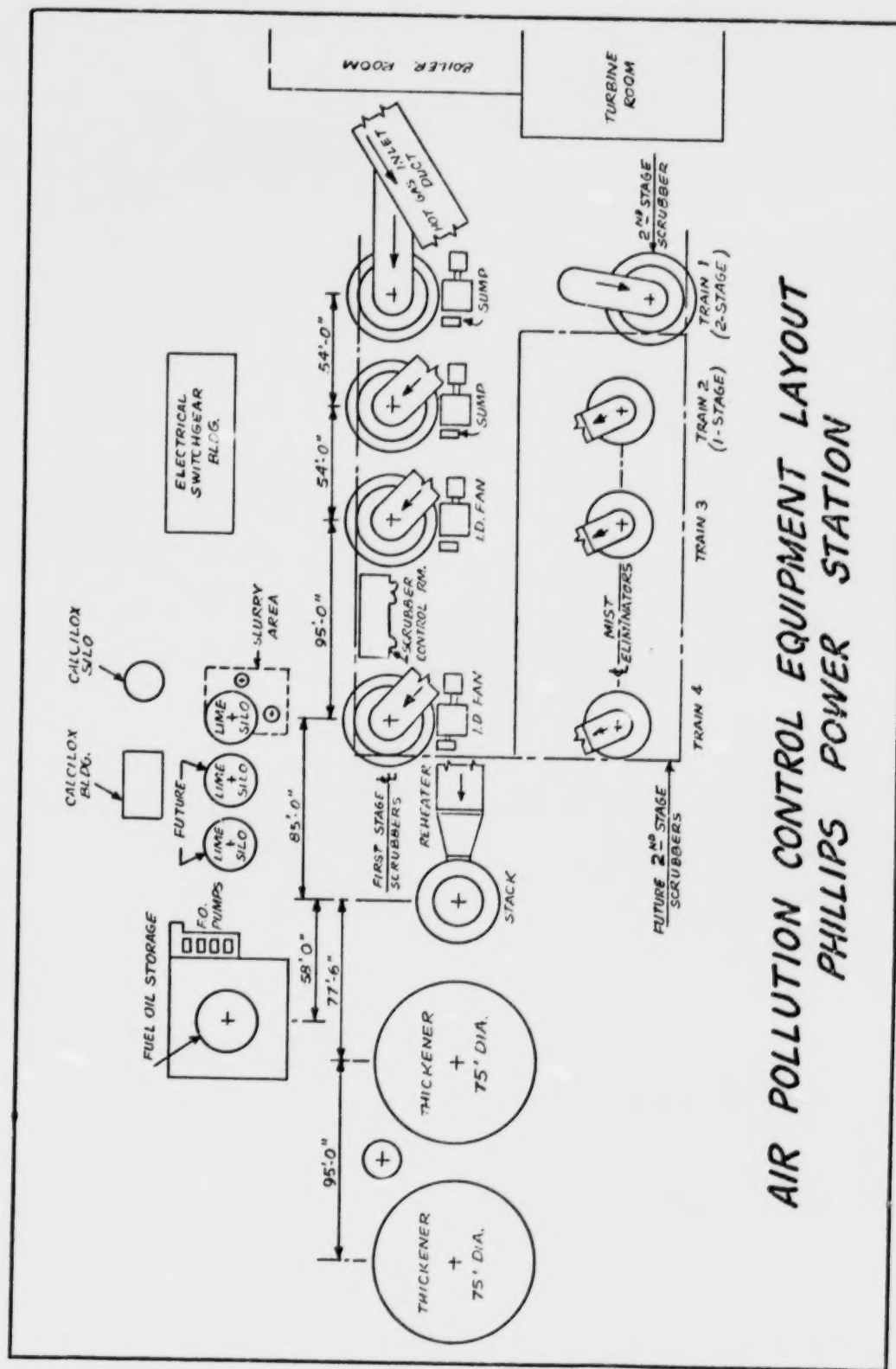
Future Objectives

Although we feel we may be able to overcome the present operating and equipment problems with additional time, effort and expense, there are five levels of performance which must be satisfactorily resolved if this flue gas desulfurization system is to be operationally feasible and economically acceptable:

1. Reliability - meaning the scrubber facility should meet the degree of availability normally expected of power generating equipment.
2. Turndown - the capability of the flue gas desulfurization equipment to follow the normal cycling operation of a power generation facility without serious disruption to both the scrubber system and the generating plant.
3. Closed Loop Operation - the ability to operate without the discharge of objectionable liquid effluents as dictated by the applicable water quality requirements.
4. Sludge - the technique for disposing of sludge from the flue gas desulfurization system without any adverse ecological effects.
5. Cost consideration - capital and operating expenses should not be of such a magnitude as to impose unreasonable financial burden on the operator/ owner and the consumer.

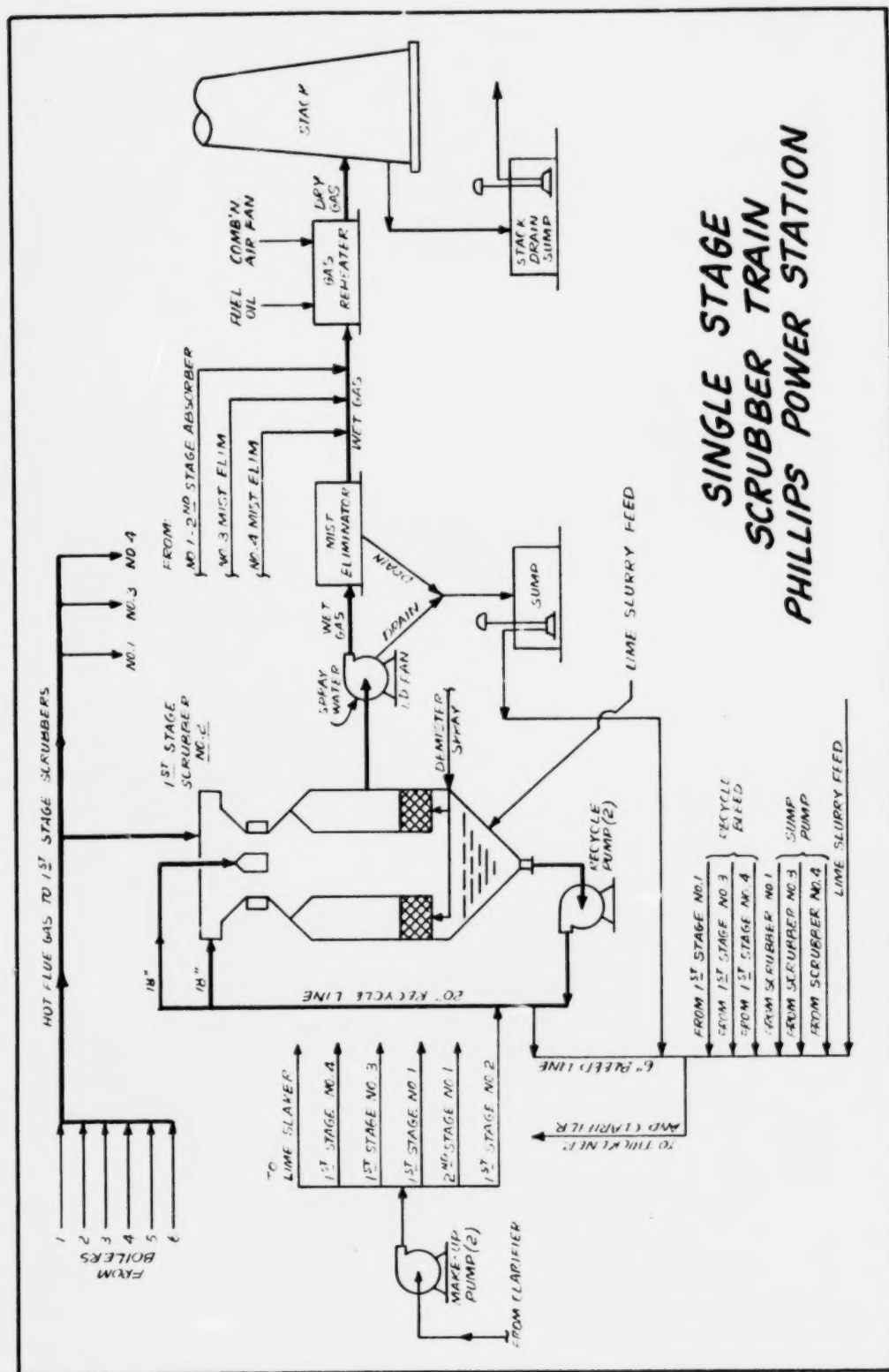
Until all these areas of concern are resolved, our system cannot be considered a successful operation as reports by others have indicated.

FIGURE 1



AIR POLLUTION CONTROL EQUIPMENT LAYOUT PHILLIPS POWER STATION

FIGURE 2



SINGLE STAGE
SCRUBBER TRAIN
PHILLIPS POWER STATION

FIGURE 3

PROTOTYPE DUAL STAGE SCRUBBER TRAIN-PHILLIPS POWER STATION

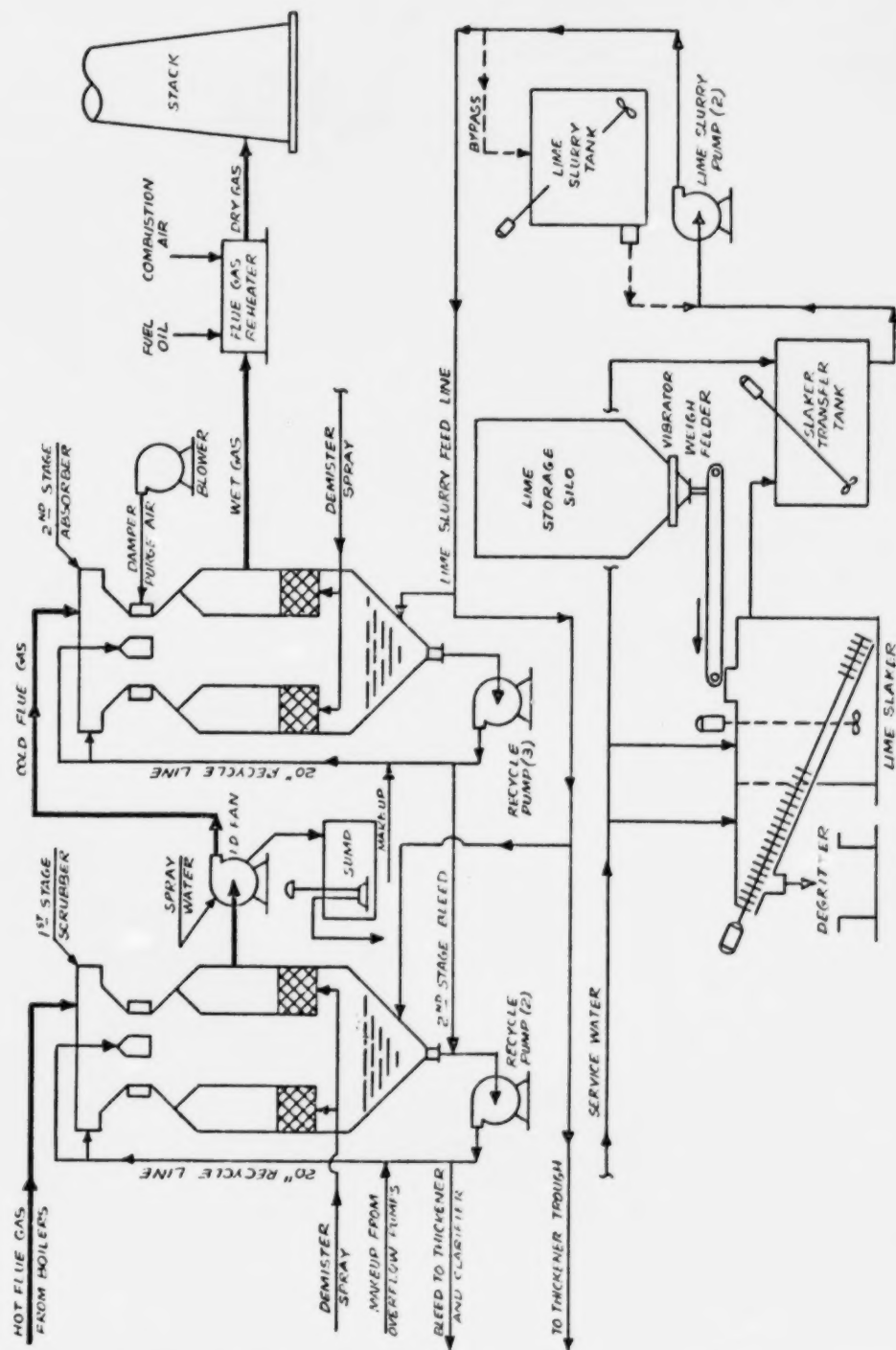


FIGURE 4

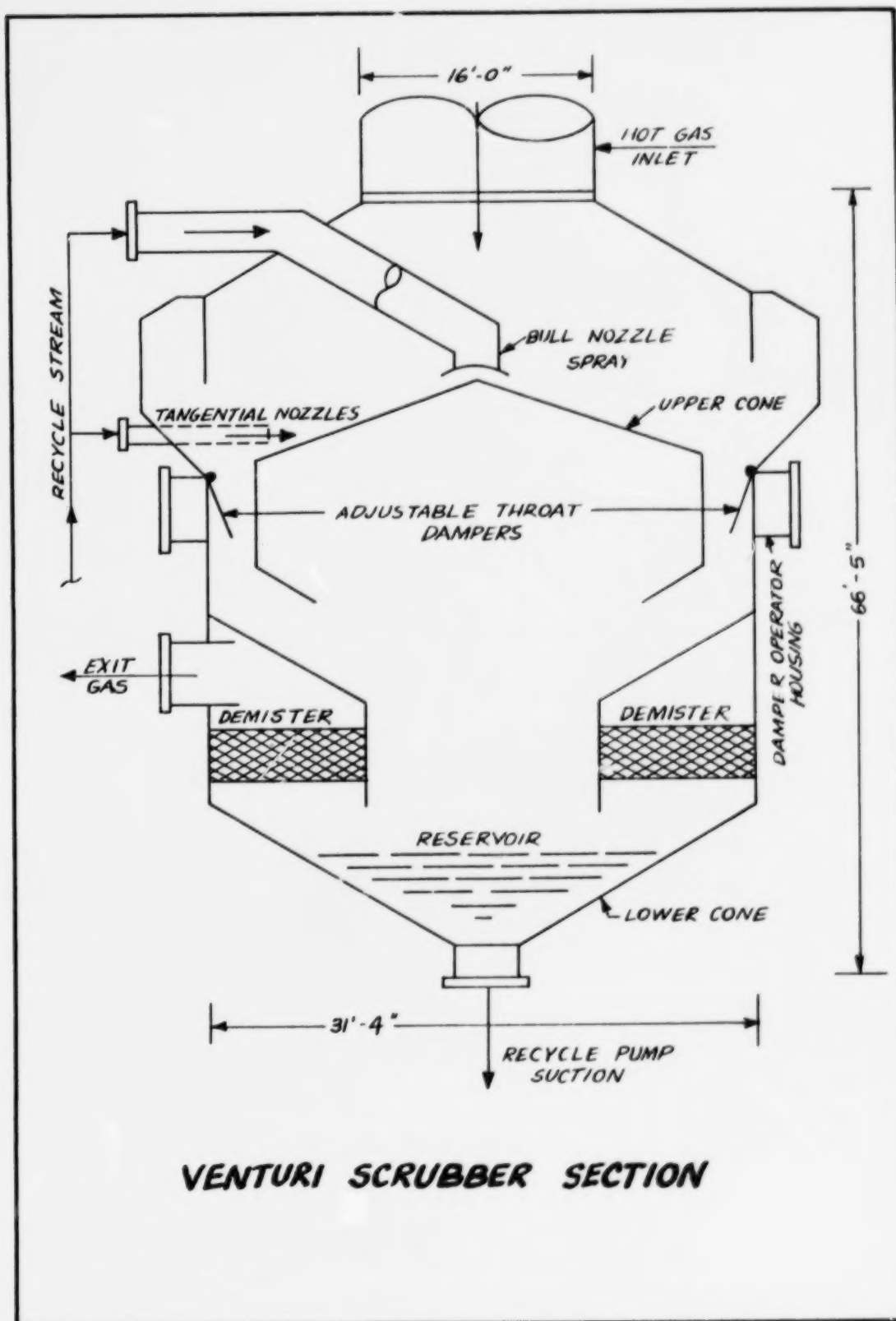


FIGURE 5
SIGNIFICANT DATES

December, 1969	-	Gibbs & Hill, Inc. requested to study methods of compliance.
September, 1970	-	Gibbs & Hill study recommends scrubber and Duquesne Light Company gave authority to Gibbs & Hill to proceed with projects at Elrama and Phillips Power Station.
October, 1970	-	Invitations were extended to bid on scrubber pilot plant.
February-May, 1971	-	Chemico pilot scrubber tests at Phillips underway.
May, 1971	-	Design of full scale system started.
July, 1971	-	Contract for scrubber process including scrubber recycle pumps, reheater and associated controls and instrumentation awarded to Chemico.
December, 1971	-	Construction started at both Phillips and Elrama plant sites.
July, 1973	-	Construction of Phase I at Phillips essentially complete. Phase I consists of one dual stage SO ₂ scrubber train and 3 single stage particulate scrubber train with associated lime and thickener equipment.
July 9, 1973	-	Operation began at Phillips with two and eventually three of station's six boilers.
October 8, 1973	-	Scrubber system shutdown for extensive repair.
March 17, 1974	-	Scrubber system began with three of six boilers.
August 26, 1974	-	Fourth boiler cut into scrubber system.
October, 1974	-	Fifth boiler cut into scrubber system.

THE HORIZONTAL CROSS FLOW SCRUBBER

by

Alexander Weir, Jr., Principal Scientist for Air Quality
John M. Johnson, Test Modules Program Manager
Dale G. Jones, Horizontal Module Project Manager
Spencer T. Carlisle, Highgrove Horizontal Scrubber Project Manager

Southern California Edison Company
for presentation at the

United States Environmental Protection Agency's
"Flue Gas Desulfurization Symposium"
Atlanta, Georgia
November 4, 1974

TABLE OF CONTENTS

	<u>Page No.</u>
Abstract	359
Foreword	360
Introduction	361
System Description	363
Operations:	369
Operations and Maintenance	369
Availability	369
Turn Down Ratio	371
pH Control	372
Closed Loop Water System	372
Performance:	
SO ₂ Removal	374
Particulate Removal	377
Lime Utilization	380
Power Consumption	383
Pressure Drop	383
Acknowledgments	386
References	387

ABSTRACT

The Horizontal Cross Flow Scrubber

The results of pilot plant testing of 9 scrubbers and the rationale for selection of two scrubber types for construction on a much larger scale at the Mohave Generating Station was described¹ at a previous EPA symposium. One of these referenced scrubbers, the TCA Scrubber, was severely damaged by a fire January 24, 1974 and is still undergoing repair at the time of this writing (August, 1974).

This paper presents the results obtained with the other large scrubber, the Horizontal Cross Flow Scrubber, since the start of its test program January 6, 1974. The results of operation at 475,000 SCFM are compared with those previously presented for operation on a much smaller scale (3000 SCFM).

Operating experience to date confirms the wisdom of the Test Module approach as the best method for achieving reliable, continuous compliance with the very stringent 0.15 lb/million BTU SO₂ removal regulations imposed by local and federal authorities.

FOREWORD

The experiments reported in this paper relating to the 1 MW and 10 MW Horizontal Scrubbers were sponsored by the Southern California Edison Company.

The 170 MW Horizontal Cross Flow Scrubber was designed, constructed and tested as part of the NAVAJO/MOHAVE TEST MODULES PROGRAM. This program was established as a joint venture of the Navajo/Mohave Power Project participants to demonstrate full-scale scrubbing as a necessary step in meeting regulatory requirements. The participants in this joint venture are:

Salt River Project Agricultural Improvement and Power District
Arizona Public Service Company
Department of Water and Power of the City of Los Angeles
Nevada Power Company
Tucson Gas and Electric Company
Bureau of Reclamation of the United States Department of the
Interior
Southern California Edison Company

Funding for this program was provided by the participants in accordance with their respective megawatt entitlements in the Navajo and Mohave Power Projects.

THE HORIZONTAL CROSS FLOW SCRUBBER

INTRODUCTION

Pilot plant testing of scrubbers was initiated at the Mohave Generating Station in July of 1971. Tests with sodium carbonate, ammonia, lime, limestone, and limestone with added soluble salts were performed with various scrubbers and combinations of scrubbers although not all reagents were tested with all scrubbers. The scrubber combinations tested on a pilot plant scale are listed.

Scrubber

1. Conventional Spray Drier Followed by a Cyclone Separator.
2. Horizontal Lime Kiln Containing Balls or Chains.
3. Single Stage Venturi with Fixed Annular Throat Followed by Centrifugal Separating Chamber.
4. Vertical Absorber Followed by Three-Stage Impingement Tray Vertical Absorber.
5. Vertical Absorber Packed with Plastic Polygrid Packing.
6. Vertical Turbulent Contact Absorber Tested with 1, 2, 3 and 4 Stages of Mobile Spheres.
7. Vertical Spray Tower.
8. Venturi Followed by Three Stage Turbulent Contact Absorber.
9. Horizontal Cross-Flow Scrubber.

Some of the results of this comparative testing has previously been reported.¹ At that time it was indicated that two of the scrubber types had been selected for further development on a larger scale than heretofore had ever been built, 450,000 SCFM each. Formal authorization was given to contractors to design and construct a Horizontal Cross Flow Scrubber system (called the Horizontal Module) and a four stage Turbulent Contact Absorber system (referred to as the Vertical Module) in December 1972 and, as previously reported, construction started on the Horizontal Module on February 15, 1973. Start-up of the Horizontal Module occurred on schedule November 1, 1974 and start-up of the Vertical Module occurred on schedule January 1, 1974. Unfortunately, a fire in the chlorobutyl rubber lining of the TCA scrubber occurred on January 24, 1974. The damage caused by this fire has still not been repaired as of this writing (September 13, 1974), and hence no operating data

on this scrubber is available and, therefore, the Vertical Module will not be discussed further in this paper. Instead, this paper will present data obtained with the full-scale Horizontal Cross Flow scrubber compared to the pilot plant data previously reported. The operating data obtained to date has justified the earlier decision to use the Test Modules approach rather than proceed with the installation of ten production modules simultaneously at the Mohave Generating Station.

SYSTEM DESCRIPTION

The Horizontal Module was built essentially as described in Figure 17 of the previous paper¹, which is presented here as Figure 1 for convenience. A simplified flow schematic is included as Figure 2. Photographs of the scrubber and some of the supporting systems are shown as Figures 3 and 4. Referring again to Figure 1, some idea of the size can be obtained by comparing the 3,000 SCFM pilot plant (and the 6 foot tall operator) shown on the left of the figure with the Horizontal Module shown on the right. Spray headers located at the top of each scrubbing chamber are 33 feet above grade. The scrubbing chamber is 15 feet high and 28 feet wide. Flue gas is drawn from the Mohave Unit 2 electrostatic precipitator outlet duct by the 1750 hp booster fan and is pumped through the scrubbing chamber and a demister section in that order. The scrubbed flue gas then is mixed with heated outside air in an indirect reheater before re-entering the precipitator outlet duct just upstream of the 500 foot stack where it is discharged to the atmosphere.

The design of the indirect reheat section contributes to the overall reliability of the scrubbing system when compared to direct reheat systems. In this system outside air provided by a 250 hp reheat fan is passed over carbon steel steam coils and the heated air is then mixed with the water-saturated flue gas. The steam coils are thus not exposed to gas more corrosive than air, and reheater performance is independent of demister performance (with direct reheater tubes, an inefficient demister would allow entrained scrubber slurry to build up on the reheater tubes, increasing flue gas pressure drop as well as accelerating corrosion of the reheater tubes themselves). Downtime for reheat cleanup and repair is thus avoided.

The remainder of the scrubbing system was also designed with reliability in mind. Since the scrubber does not contain any internal packing or trays, buildup and plugging in the scrubber itself with fly ash/scrubber slurry mixture does not occur, either during operation or on shutdown. During operation the lime scrubbing slurry is pumped in a counter-current manner from the bottom of one stage to the top of the next stage. Three slurry pumps are used for each stage and, as indicated in Figure 2, they are located at the bottom of pyramidal hoppers. The use of three pumps allows the benefits of redundancy since at nominal L/G ratios only two pumps are required. Satisfactory operation has been obtained with only one pump per stage, thus allowing pump maintenance to be performed without scrubber shutdown.

While four stages of scrubbing are used, the scrubber can continue to operate with three, two or one stages of scrubbing, although as may later be seen with considerable degradation of SO₂ removal performance (only 70% SO₂ removal with two stages). This

FIGURE 1

**SIZE COMPARISON BETWEEN PILOT
AND FULL-SCALE HORIZONTAL SCRUBBERS**



PILOT FACILITY

FULL-SIZE HORIZONTAL TEST MODULE

[illegible]

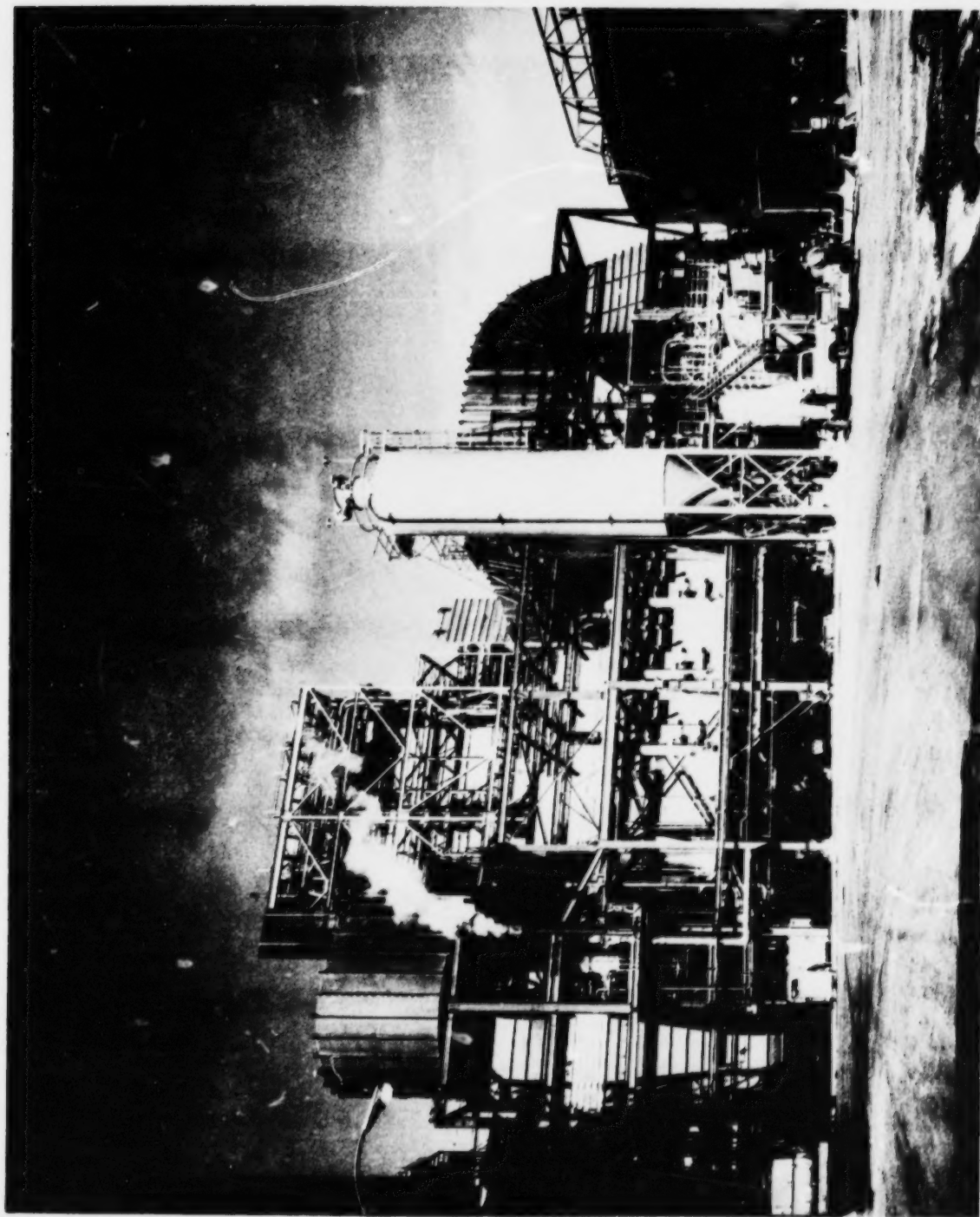
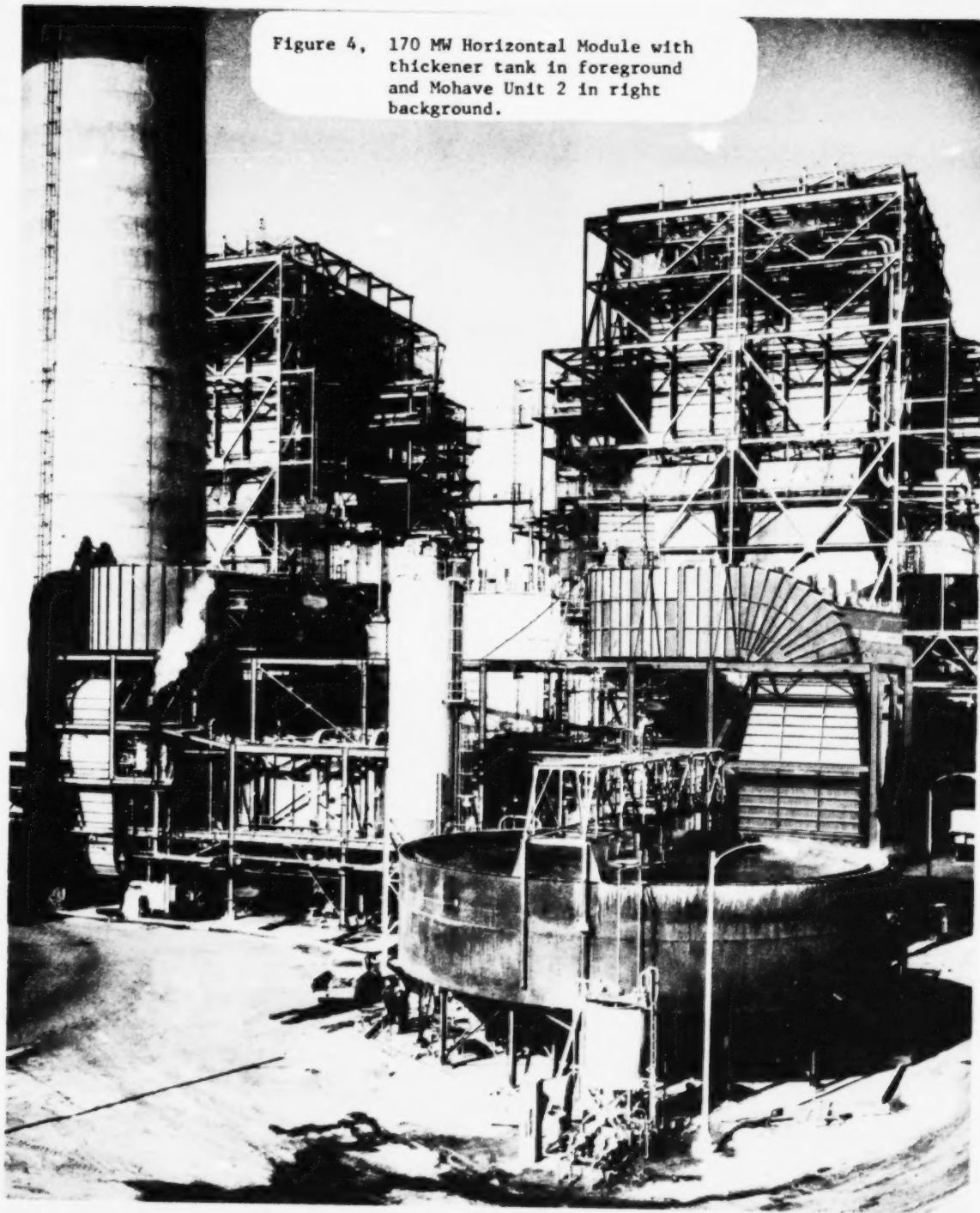


Figure 3, 170 MW Horizontal Module with
lime silo in background

Figure 4, 170 MW Horizontal Module with thickener tank in foreground and Mohave Unit 2 in right background.



redundancy allows scrubber maintenance or repair problems to be postponed until the generating unit is shut down.

To digress for a moment, the modular approach adopted for the Mohave Generating Station also includes the benefits of redundancy. As previously indicated, the Mohave participants decided to test full scale modules before proceeding with the installation of 10 or 12 scrubbers. Thus, operating problems encountered could be solved once, rather than trying to fix the same problem simultaneously on a number of scrubbers. While a selection of scrubber types has not been made, present plans are either to install five vertical modules as presently designed plus one spare per generating unit or four Horizontal Modules as presently tested plus one spare per generating unit. Thus an individual module availability would only have to be 83.3% in the case of the TCA scrubber or 80.0% in the case of the Horizontal Scrubber to allow 100% scrubber availability to the generating unit operating at full load. It is believed that the "elasticity" incorporated in the Horizontal Module will allow it to meet that goal.

OPERATIONS

Operations and Maintenance

The test program has been scheduled to operate the scrubber continuously for 24 hours a day, 7 days a week in order to obtain the maximum amount of operating time and data.

All scrubber operations are performed by SCE plant equipment operators (PEO's) and operating foremen that are also rotated to other station activities. One Generating Station "Watch" engineer has been assigned full time as the operating supervisor for both the Vertical and Horizontal test programs. SCE chemical and instrument technicians are available as required. Although many of the test conditions that are peculiar to the test program sometimes make operations more complex than would normally be expected, it has been demonstrated that two PEO's and one foreman, per shift, can operate the Horizontal Module. Maintenance support is contracted and consists of one millwright and one pipefitter during the day shift, 5 days per week, and one laborer during the day shift, 7 days per week.

For both test programs three SCE Research and Development engineers have been assigned at Mohave for test planning, test monitoring, data analysis, evaluation and reporting. Special chemical laboratory support and particulate/gas sampling tasks have been contracted.

It is anticipated that the operation of multiple commercial type Horizontal Scrubbers would be a relatively simple task that can be handled by a minimum number of PEO's who have completed the standard apprentice training program.

Availability

The Horizontal Test Module at Mohave receives flue gas only from Unit 2 and does not have the inlet ducting cross-connected to Unit 1. As a result, the scrubber is only operated when Unit 2 is in service, i.e., when the unit is down for any reason the scrubber is shutdown. Therefore, availability has been calculated with respect to Unit 2 operating time as opposed to calendar time or other reference standards. Since the start of the test program on January 16, 1974 and as of September 13, 1974 Unit 2 has been in-service for a total of 5,003 hours and the scrubber in operation for a total of 4,292 hours resulting in an overall availability of $4292/5003 \times 100 = 85.4\%$. This percentage is the accumulative total availability and does not distinguish between scrubber outages that were scheduled for inspections and/or test program configuration changes, those that were necessitated because of test module problems

or unit outages which allowed for "free" maintenance time on the scrubber. A brief description of both categories are listed below:

Planned or Scheduled Scrubber Outages = 297 Hours

	<u>Hours</u>
Inspect all spray nozzles	19
Install full compliment of spray nozzles for high L/G tests	92
Remove spray nozzles to revert to design configuration	98
Inspect demister blades	4
Hardware configuration change and maintenance	<u>84</u>
Total	297

Test Module Problem Outages = 414 Hours

1. Clean mix tank screen - 4 hours.

Problem: Overflow of the 48,000 gallon mix tank.

Cause: Debris from the first stage hopper clogged the mix tank inlet screen.

Correction: Replaced the 1/4" mesh screen with an expanded metal screen containing 3/4" openings.

2. Repair and clean slaker/piping - 41 hours.

Problem: Loss of lime feed slurry to the mix tank.

Cause: Lime slaker and associated piping was plugged with hard calcium sulfate scale. Formation of this scale resulted from using scrubber process water for slaking.

Correction: Converted to use of station service water for feeding the slaker, transfer pump seals and lime slurry feed tank level control bubblers. Also replaced major portions of the slaker related carbon steel piping with rubber hose and eliminated most all of the horizontal pipe runs and 90° elbows.

3. Remove two hard hats from the thickener - 94 hours.

Problem: Plugged thickener tank outlet.

Cause: Excessive winds, up to 45 MPH, caused two operators working at the thickener to lose their hard hats. The hats lodged in the thickener outlet drain cone above the discharge valve.

Correction: Two scuba divers, in wet suits, were eventually able to retrieve both hard hats but only after the bed and liquid levels of the thickener were lowered. A storage box for hard hats has been placed at the bottom of the thickener tank ladder.

4. Repair inlet isolation damper - 70 hours.

Problem: Inlet isolation damper would not close.

Cause: Both push rod shafts were bent due to operation with improperly adjusted jack screws.

Corrections: Replaced both push rods, added thrust bearings to the jack screws and installed heavy duty limit switches.

5. Align booster fan/motor shafts - 167 hours.

Problem: Excessive vibration in booster fan/motor shafts.

Cause: Gradual increase in vibration during the early summer high temperatures at Mohave. Motor and concrete foundation were placed during cooler weather during the fall of 1973 and expanded with the temperature increase.

Correction: Realigned the shafts and measured the alignment numerous times to obtain and confirm the proper settings.

6. Repair inlet duct turning vanes and scrubber hopper flange leaks - 38 hours.

Fatigue failure which caused portions of a turning vane to fall off and corrosion failure of two carbon steel hopper flanges required repair work.

Turn Down Ratio

The Horizontal Module was designed for a nominal operating inlet gas flow rate of 450,000 SCFM. This corresponds to a superficial linear velocity of 21.6 ft/sec in the scrubber shell, based on the

1 MW pilot plant tests. Results of flexibility testing indicate the scrubber can be operated satisfactorily over a wide range of inlet flue gas flow rates. The highest gas flow obtained to date has been 544,000 SCFM (SO₂ removal of approximately 94%) and represents the upper limit of the booster fan capacity. The lowest gas flow rate tested has been 130,000 SCFM (97.4% SO₂ removal) and represents what is believed to be equivalent to one-fifth of the lowest gas flow rate (or load) at which one Mohave boiler unit would be operated.

pH Control

The lime slurry feed rate to the reaction mix tank is automatically controlled to maintain a constant slurry pH. As anticipated, this automatic pH control method provides a very stable and finely tuned chemical control system for the scrubbing system. Results from the long duration material balance tests indicate that the control system ensures both a high SO₂ removal and a high utilization of the added lime under steady state conditions. Results of the shorter flexibility tests have demonstrated the pH control system capable of "following" fluctuations in flue gas flow rate and inlet SO₂ concentrations.

Closed Loop Water System

The front and backend systems are integrated with the scrubber to provide a closed loop water system. This allows the makeup water flowrate to be a function of flue gas flowrate (and SO₂ content) by providing all internal water requirements with recycled water. The scrubbing system can thus be turned down to any desired capacity while simultaneously turning down the makeup water flowrate to avoid operating the system out of water balance.

The makeup requirement at a flue gas flowrate of 440,000 SCFM and 280°F (averaged over 1036 hours of continuous testing) is 152 gpm. 82% of this total was cooling tower blowdown water and 18% was service water required for slaking and dilution of lime slurry. 93% of the makeup water goes up the outlet duct as water vapor and the remainder is water of hydration, interstitial water and evaporation which cannot be reclaimed from the waste sludge disposal pond. The scrubber area is paved with concrete and provided with curbing. All seal water drips, hose washdown water, spills and pond supernatant water are returned to the scrubber system to maximize waste water utilization. No water is put into the ground or the station peripheral drainage system.

The most critical part of the water recycle concept required for closed loop operation is the source of pump and instrument seal water. The seal water requirement represents a flowrate of 50 gpm, which is independent of flue gas flowrate or unit load.

With high levels of dissolved solids in the makeup water, the total dissolved solids in the scrubber slurry reach extremely high levels - 15 to 20 weight percent. The salts represented are largely sodium chloride (NaCl), magnesium sulfate (MgSO_4) and sodium sulfate (Na_2SO_4).

Thickener overflow is filtered to provide the pump and instrument seal water. In spite of the high dissolved solids no severe chemical scaling or corrosion has been observed.

PERFORMANCE

SO₂ Removal

The SO₂ removal results of the full-scale Horizontal Scrubber are compared in Figure 5 with data from the original 1 MW pilot plant scrubber at the Mohave Station. For both the pilot plant and the full-scale test module, 90% SO₂ removal was achieved at a liquid to gas contacting ratio (L/G) of about 17.5 gallons per 1000 SCF, using four stages of scrubbing. The degree of repeatability between pilot scale results and full-scale results is noteworthy. The performance of the full-scale test module is slightly better than the pilot plant at reduced L/G ratios.

The criterion for the Mohave Station to be in compliance with the Clark County SO₂ regulations is equivalent to about 50 ppm at full load. The scrubber design criterion was established at an outlet SO₂ concentration of 40 ppm. This criterion was met or exceeded by the full-scale Horizontal Scrubber at 200 ppm inlet SO₂ concentration with an L/G ratio (per stage) as low as 12.5 gallons per 1000 SCF.

Percent SO₂ removal at various L/G ratios is shown plotted for three different gas flow rates in Figure 6. These results show more efficient SO₂ removal at lower gas flow rates (and increased residence time in the scrubber) as would be expected. As mentioned earlier, the Horizontal Module can be operated with any number of stages. Figure 7 shows SO₂ removal efficiency as a function of the number of stages and indicates that 90% SO₂ removal can be achieved with three scrubbing stages at the specified conditions.

The sulfur content of the coal burned at Mohave has averaged 0.38% which results in an average stack gas SO₂ concentration of 200 ppm. A recent coal survey of the Black Mesa mine field indicates the highest coal sulfur content ever to be expected will be 0.83% which corresponds to approximately 400 ppm SO₂. Most of the original pilot plant work was performed at this level and future plans are to inject gaseous SO₂ to achieve this level for testing the Horizontal and Vertical Modules. A good estimation of the performance expected from the Horizontal Module at 400 ppm inlet SO₂ (and higher) can be obtained from the results of SCE's Highgrove Generating Station 10 MW Horizontal Scrubber Pilot Plant Experiment ("to test the feasibility of scrubbing high sulfur oil-fired flue gas to the equivalent of low sulfur fuel oil"). The Highgrove 10 MW scrubber design was based on results from the high inlet SO₂ experiments performed on the Mohave 1 MW pilot unit, and is essentially identical to the 450,000 SCFM Horizontal Test Module with respect to operating parameters.

Figure 8 presents a comparison of outlet SO₂ concentration vs. L/G ratio of the 1 MW and 10 MW scrubbers at an inlet SO₂

FIGURE 5
SO₂ REMOVAL VS L/G RATIO
COMPARISON 1MW PILOT & 170 MW MODULE

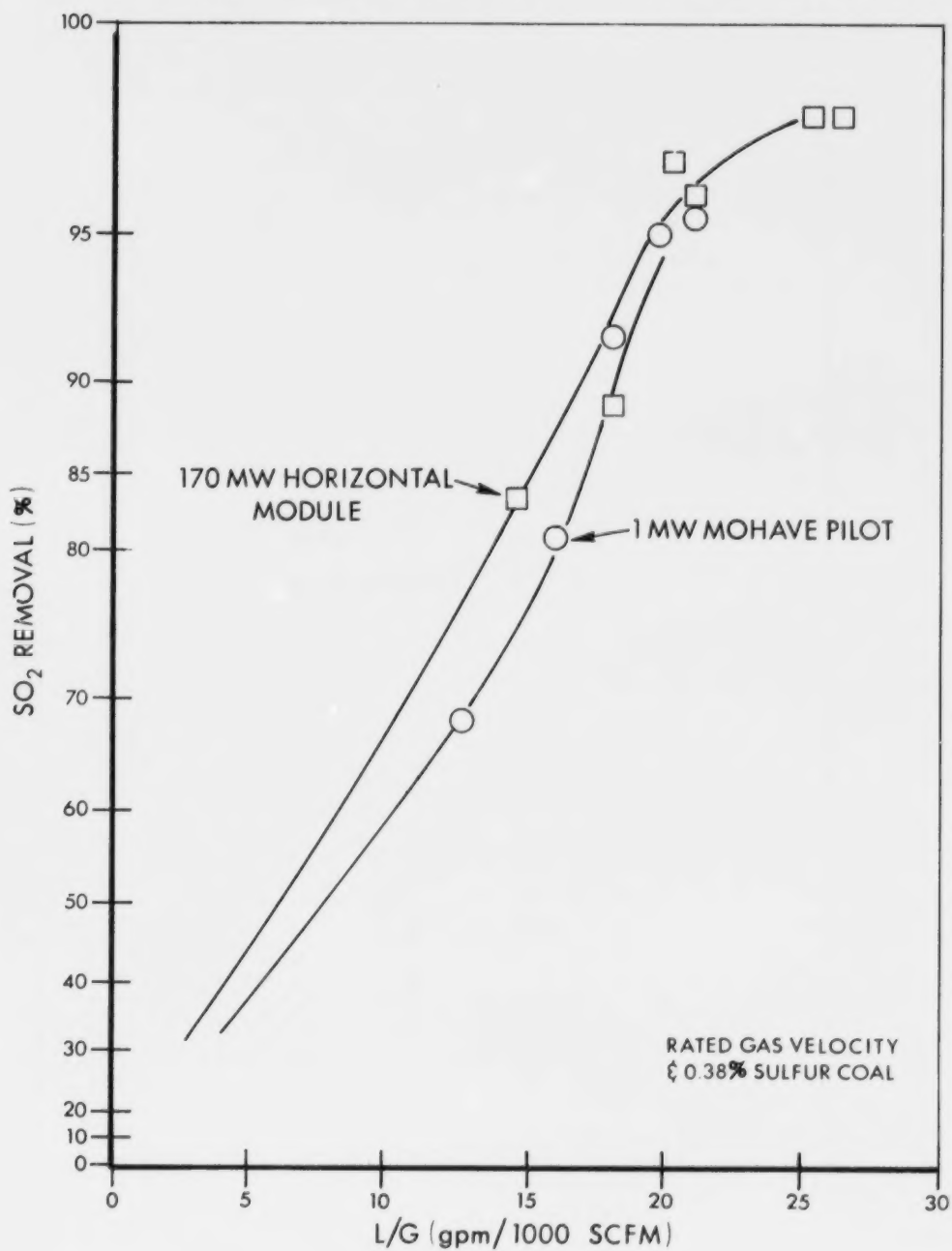
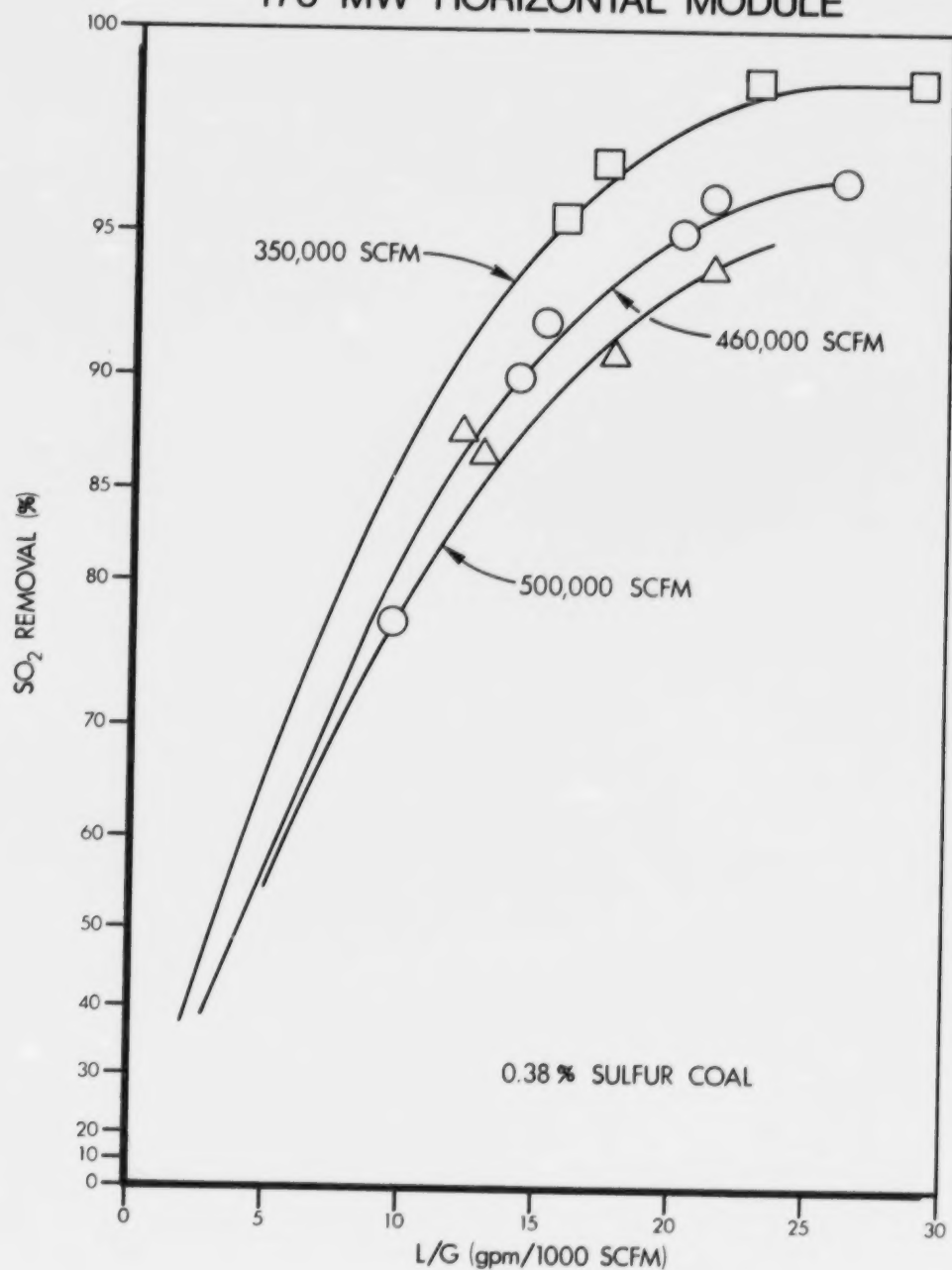


FIGURE 6

SO₂ REMOVAL vs L/G RATIO 170 MW HORIZONTAL MODULE



concentration of 400 ppm. Again, the degree of repeatability between the pilot and the 10 MW version is quite good. From this plot we can reasonably expect the 450,000 SCFM Horizontal Module to meet the 90% removal criteria (at 400 ppm inlet SO_2) with an L/G ratio (per stage) of approximately 17.5 gallons/1000 SCF.

Figure 8 also presents the curves for the pilot plant and 10 MW scrubbers at the 1000 ppm inlet SO_2 level. It is obvious that the 10 MW Highgrove scrubber performs much better than expected with respect to L/G ratio. The lower L/G ratio requirement for the Highgrove scrubber to achieve the same degree of removal as the pilot plant was obtained by optimizing the placement and number of nozzles in the scrubber chamber.

Experiments with the pilot plant scrubber indicated that inlet SO_2 concentration was an important variable in SO_2 removal. Pilot plant test results indicated a maximum SO_2 removal efficiency under fixed operating conditions at 400 ppm inlet and a rapid decrease in removal efficiency with increases in inlet SO_2 concentration above 400 ppm. The effect of decreasing removal efficiency above 400 ppm was not as pronounced with the 10 MW scrubber as shown in Figure 9. For the Highgrove scrubber the slight reduction in SO_2 removal efficiency above 1600 ppm SO_2 can be minimized by either a small increase in the L/G ratio or by the addition of a fifth scrubbing stage. Recent tests have been made at Highgrove utilizing five stages of scrubbing at inlet concentrations as high as 3000 ppm.

Particulate Removal

An objective of the Horizontal Scrubber Test Program equal in importance to SO_2 removal has been that of particulate removal downstream of the 98% efficient, cold-side, electrostatic precipitators at the Mohave Station. Since the ash content of the coal is nominally 9%, the grain loading downstream from the precipitator is nominally 0.07 gr/SCF, but varies with coal quality, unit load, and operating conditions from 0.01 gr/SCF to perhaps 0.10 gr/SCF.

Previous pilot plant results with the 3000 SCFM Horizontal Scrubber indicated about 90% particulate removal at representative grain loading conditions. This result has been re-confirmed with the full-scale Horizontal Module, operating above its design conditions at a flue gas flowrate of 474,000 SCFM.

Inlet and outlet particle grain loadings under constant scrubber operating conditions are presented over a wide range of inlet grain loadings in Figure 10. As shown, the particulate removal varies with inlet grain loading from 70% removal at 0.01 gr/SCF to 98% removal at 1.00 gr/SCF.

FIGURE 7
SO₂ REMOVAL vs. No. OF SPRAY STAGES
170 MW HORIZONTAL MODULE

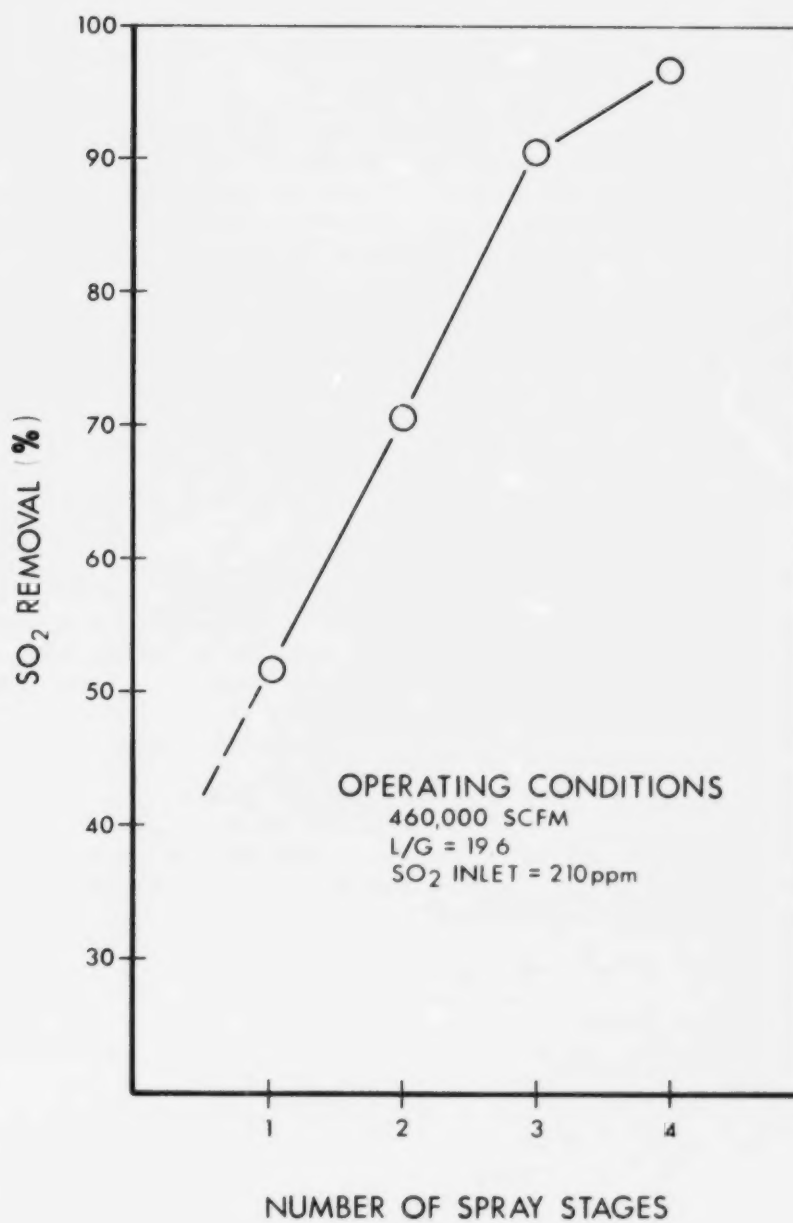
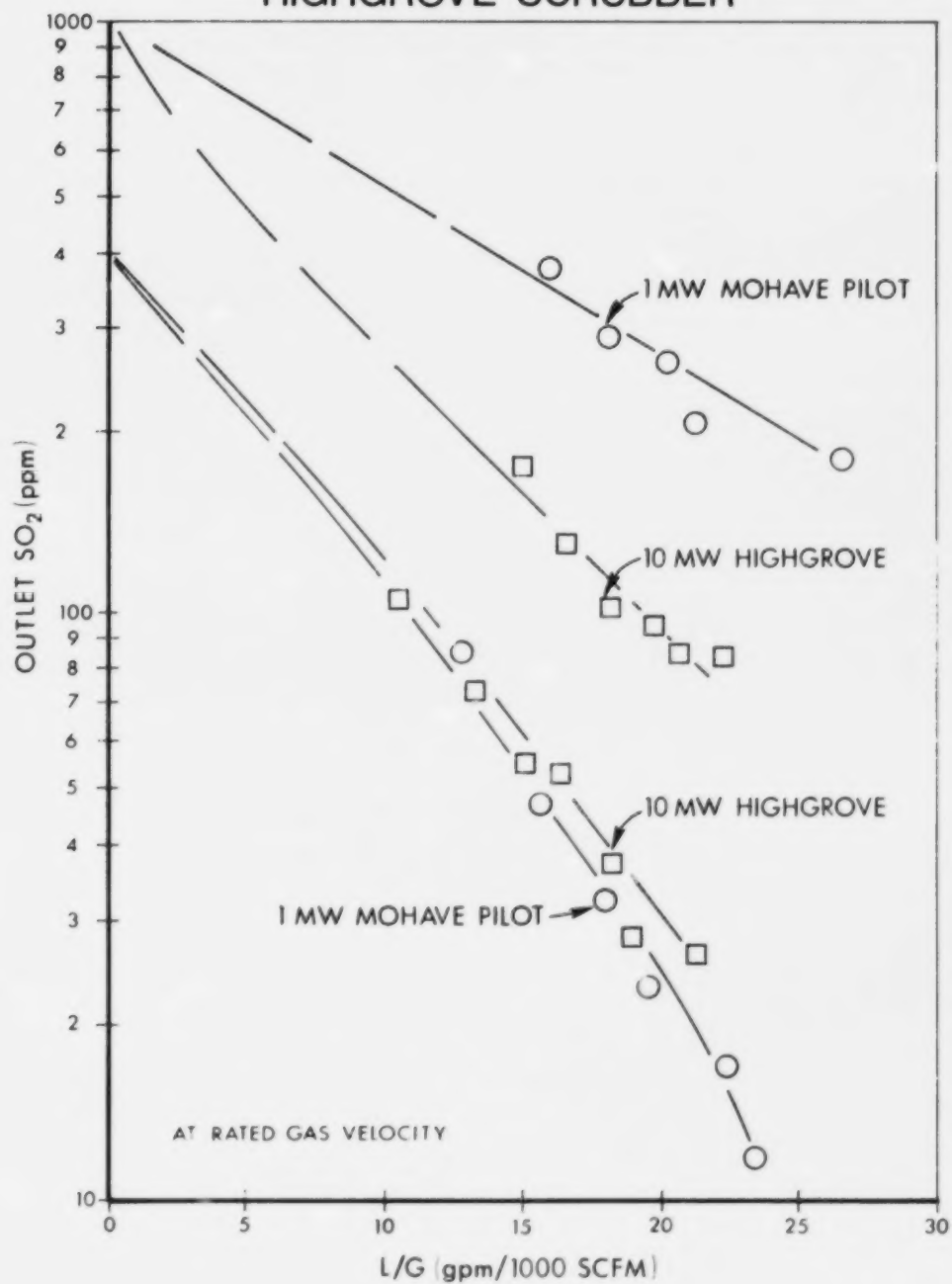


FIGURE 8
 OUTLET SO_2 vs. L/G COMPARISONS-
 1 MW MOHAVE PILOT AND 10 MW
 HIGHGROVE SCRUBBER



The increased particulate removal efficiency at higher inlet grain loadings is undoubtedly due to the fact that at higher dust loadings the stack gas is composed of larger diameter particles due to precipitator sections being out of service.

Particle size in the flue gas at Mohave are normally very small. About 90 cumulative wt % of the particles are less than 4 microns in diameter with 70 wt % less than 2.0 microns in diameter, 40 wt % less than 1.0 micron in diameter and about 15 wt % less than 0.5 microns in diameter.

Fractional size collection efficiencies previously reported in last year's EPA symposium for the pilot plant are compared below with full-scale test results.

<u>Particle Size</u>	<u>3000 SCFM</u>	<u>474,000 SCFM</u>
	<u>Horizontal Pilot Plant</u>	<u>Horizontal Scrubber</u>
	<u>at an inlet grain</u>	<u>at an inlet grain</u>
	<u>loading of 0.02 gr/SCF</u>	<u>loading of 0.08 gr/SCF</u>
Greater than		
1.5 microns	97%	92%
1.0 to 1.5		
microns	96%	91%
0.5 to 1.0		
microns	87%	85%
0.3 to 0.5		
microns	75%	76%

Lime Utilization

High calcium pebble lime delivered from Nevada indicated the following average composition:

CaO alkalinity as CaO (CaO, Ca(OH) ₂ , and CaCO ₃)	95.3%
Total alkalinity as CaO (CaO, MgO, Ca(OH) ₂ , CaCO ₃ and other trace alkaline materials	97.4%

Lime utilization was calculated by three different methods over an operating period from May 6 through May 25, 1974. Other sources of alkalinity entering the scrubber system were also taken into account. Chemical analysis showed that flyash and scrubber makeup

FIGURE 9

EFFECT OF INLET SO_2 CONCENTRATION ON PERCENT SO_2 REMOVAL

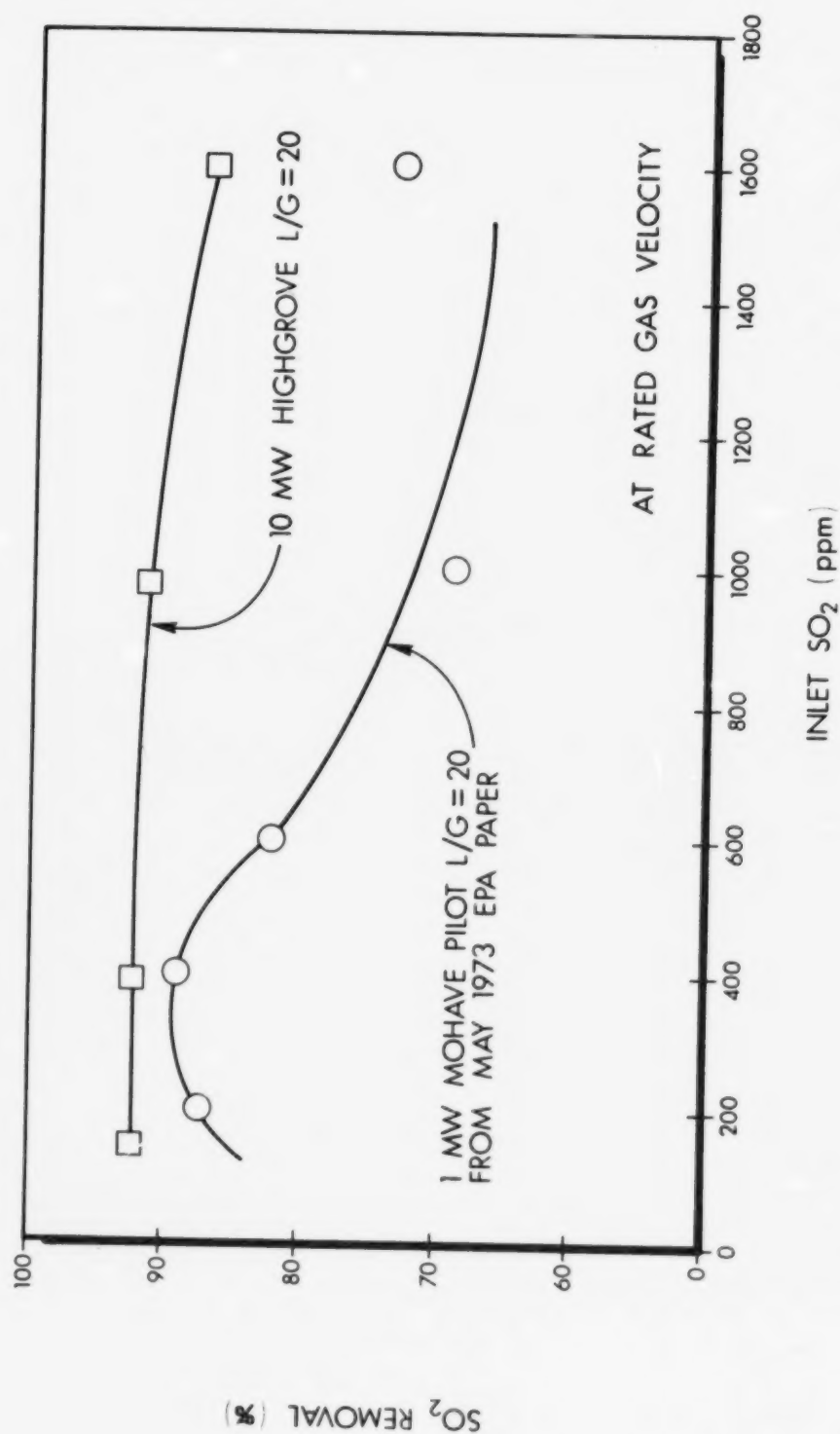
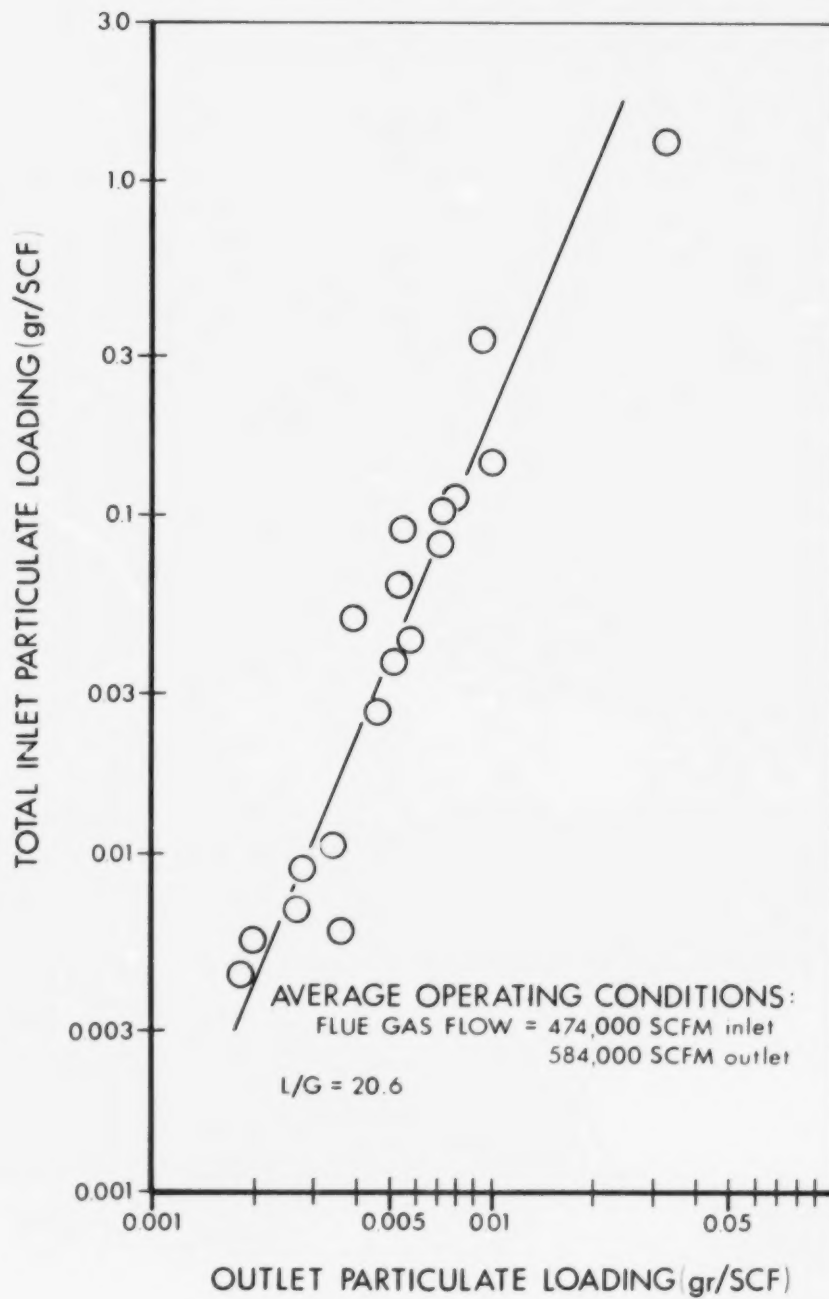


FIGURE 10
INLET vs. OUTLET GRAIN LOADING
170 MW HORIZONTAL MODULE



water contributed respectively 1.17% and 0.85% of the equivalent CaO fed to the system. The three lime utilization determinations are summarized as follows:

Method A: CaO utilization (from SO₂ removed from gas) = 99.75%

Method B: CaO utilization (from unreacted CaO in sludge and material flowrates) = 99.40%

Method C: CaO utilization (from chemical analysis of sludge independent of material flowrates) = 99.27%

The average utilization for the three methods is 99.47% which is nearly stoichiometric.

Power Consumption

The electric power consumed by the Horizontal Scrubber is metered at the 4160 volt bus and includes transformer losses for voltage less than 4160 volts, control room power and air conditioning, scrubber lighting and miscellaneous welding and maintenance power usage. Not included are the following usages of power:

- A. Reheat steam at 40,000 lb/hr, 400°F and 350 psig
- B. Service air at 40 SCFM, 90 psig
- C. Electric power to operate sludge stabilization mixing equipment (4 kW)

The electric power usage was monitored during 556 hours of continuous operation. The average flue gas flowrate during this period was 451,000 SCFM. The scrubber power consumption averaged 2.66 MW. In general, flue gas flowrate per MW varies from station to station. For the 790 MW Unit Two at McHove, the flue gas flowrate at full load is between 1.9 and 2.1 X 10⁶ SCFM (at 600°F and 1 atm). Thus a flowrate of 451,000 SCFM is equivalent to a range between 170 and 188 MW, and the power consumption measured corresponds to a range between 1.42 and 1.57% of the power generated.

Pressure Drop

One important characteristic of a scrubber which contributes a great deal to both the capital cost and the operating cost is the pressure drop through the scrubber. The pressure drop is, of course, primarily a function of the design of the scrubber.

As discussed in last year's EPA symposium, the Horizontal pilot plant was characterized by a low pressure drop compared with the other

pilot plants. Specifically Figure 14 of Ref. 1 indicated that the pressure drop across the 3000 SCFM Pilot Plant Horizontal Scrubber itself was 1.0 inches of water while Figure 19 indicated that the total system pressure drop (at nominal conditions) anticipated for a 450,000 SCFM Module System was 6 inches of water.

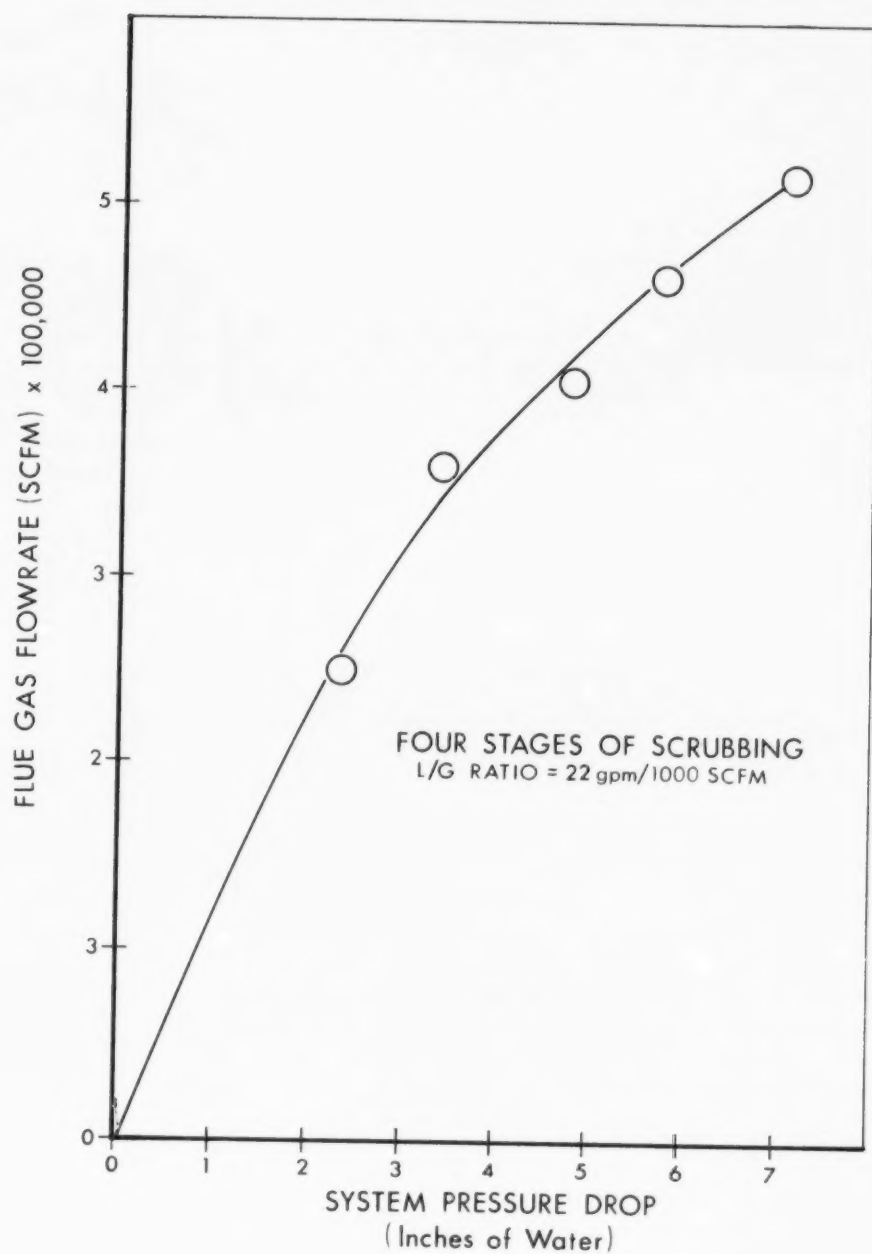
In Figure 11, the total system flue gas pressure drop is plotted versus gas flowrate at a constant L/G ratio. As may be seen, the total system pressure drop at 450,000 SCFM is 5.5 inches of water. This may be compared to the 6 inches of water previously predicted. The total system pressure drop is the differential pressure developed by the booster fan as required to draw flue gas from the precipitator outlet duct and drive it through the scrubber, demister, reheat mix chamber and outlet ductwork into the stack breeching duct.

The test data in Figure 11 indicates that the total system pressure drop under design operating conditions was approximately 6.0 inches of water. The various scrubber components contributing to this total are roughly as follows:

<u>Component</u>	<u>Differential Pressure Inches of Water</u>
Inlet Ductwork	1.0
Scrubbing Chamber	1.0
One Demisting Stage	0.5
Indirect Reheating Chamber	2.5
Outlet Ductwork	<u>1.0</u>
Total	6.0

The pressure drop through the indirect reheating chamber is somewhat high to insure complete mixing of hot and cold gas streams.

FIGURE 11
SYSTEM PRESSURE DROP vs.
FLUE GAS FLOWRATE
170 MW HORIZONTAL MODULE



ACKNOWLEDGMENTS

The experiments reported in this paper relating to the 1 MW and 10 MW scrubbers were sponsored by the Southern California Edison Company. Construction and Testing of the 170 MW scrubber was funded by the Mohave Project and Navajo Project participants - Arizona Public Service Company, City of Los Angeles Department of Water and Power, Nevada Power Company, the Salt River Project, Southern California Edison Company, Tucson Gas and Electric Company and the United States Department of the Interior, Bureau of Reclamation with costs allocated in accordance with their megawatt entitlements in the Mohave Project and in the Navajo Project. The Salt River Project serves as Project Manager for the Navajo Project while Southern California Edison serves as Project Manager of the Mohave Project as well as Project Manager for the Test Modules Program.

The design, construction and operation of the Horizontal Module required the efforts of many individuals and organizations. The courageous management decision to build the Horizontal Module 150 times larger than the pilot plant was made by Mr. Tom Morong, Chief Engineer and Assistant General Manager of the Salt River Project and Mr. Jack B. Moore, Vice President of the Southern California Edison Company. Mr. Richard Durning of the Salt River Project made many contributions, particularly in the design criteria and test planning area.

Stearns Roger, Inc. performed the engineering design according to SCE design criteria and also procured the major equipment and provided an engineer for test support. The Bechtel Power Corporation performed all construction activity at the Mohave Generating Station. Routine maintenance and design modification services have been performed during the Test Program by Jesco, Inc. Truesdail Laboratories has also supported the test program with field chemists, and gas sampling crews.

Many Edison personnel have contributed a great deal to the success of this program; these include Messrs. B. C. Stowers and J. Harvey of Construction Engineering, Mr. N. J. Dellaven of Generation Engineering, Messrs T. L. Reed, R. L. Manley and C. Patel of Cost Schedule Engineering, and Messrs. H. A. Kerry, E. A. Danko, S. D. Sharp, W. C. Martin and R. B. Rolfe of Edison's R&D Organization. Mr. G. L. Fraser, Station Superintendent and Mr. R. Young of the Mohave Generating Station Staff have successfully supervised the operation and maintenance of the scrubbers. The contributions of the SCE operators and foremen themselves cannot be overstated.

Finally, Dr. L. T. Papay, Director of Research and Development for Southern California Edison Company has served as Chairman of the Coordinating Committee for the Navajo/Mohave Test Modules Program. His encouragement and technical contributions are sincerely appreciated.

REFERENCES

Alexander Weir, Jr. and Lawrence T. Papay "Scrubbing Experiments at the Mohave Generating Station" U. S. Environmental Protection Agency's Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14, 1973.

OPERATIONAL STATUS AND PERFORMANCE OF THE LOUISVILLE
FGD SYSTEM AT THE PADDY'S RUN STATION

Mr. Robert P. Van Ness, Manager Environmental Affairs
Louisville Gas And Electric Company

November 4 - 7, 1974
EPA, FGD Symposium, Atlanta, Georgia

ABSTRACT

Operations at the Closed Loop Full Scale 65MW SO₂ System commenced in April 1973 and has operated in limited and extended modes to date. Absolute reliability of the system has not been established; however, operations to date have been extremely satisfactory and gratifying. All data suggests high attainments of SO₂ and particulate matter requirements without operational difficulties such as scaling, plugging, and corrosion or erosion problems. High degree of operational on time at varying loads with various levels of sulfur coal input. Extremely high removal rates of efficiencies of SO₂ in 1974 with inoculations of varying amounts of magnesium. Stoichiometric requirements of the system have been excellent.

OPERATIONAL STATUS AND PERFORMANCE OF THE LOUISVILLE
FGD SYSTEM AT THE PADDY'S RUN STATION

I have been requested to present to this Symposium a status report on the operation of the Paddy's Run Unit #6 SO₂ removal system to date and the current status of the compliance program proposed to meet local air pollution standards.

Our Ca(OH)₂ (carbide lime) slurry tail end wet scrubbing system has been described in detail by both Combustion Engineering Co. and Louisville Gas & Electric Co. in previous statements and in publications. So, in the interest of avoiding repetition, I will assume that everyone is basically familiar with the installation. If not and more information is desired, I would recommend that you obtain a copy of a recent paper presented by Messrs. Martin, Plumley and Minor at the National Coal Association Symposium on "Coal and the Environment" in Louisville, Ky. on October 23. The paper entitled "The C. E. Lime Wet Scrubbing Process from Concept to Commercial Operation" covers in considerable depth the data, concepts, flow conditions and charts pertaining to the system and its operation.

As a start, flow diagrams of this system are enclosed (Exhibit A), for your edification which basically shows the equipment and overall design flows and enumerates the stages of control of the calcium bisulfite and calcium sulfite reactions using a waste product known as carbide lime additive. Basically, the concept has been to remove the SO₂ from the flue gases by the use of a liquid to gas contactor such that the SO₂ is removed in a soluble form (calcium bisulfite) and to control the chemistry so that the solid salts of calcium sulfite and calcium sulfate are not formed in the scrubber. The former is a critical item in that it tends to come out of solution at a pH of about 6.2. The sulfate formation is an oxidation reaction which must be controlled by various techniques such as pH, time, liquid and gas flows, and seed crystals. The control concept and equipment have produced very satisfactory results.

In our opinion the overall reliability and efficiency of this SO₂ removal system to date has been good. Starting in April of last year the unit was able to operate 70% of the time that the boiler was available for service through 9 months of 1973 and an availability of 98% during the last 4 months of the year.

The low 70% availability factor was for the most part due to various modifications which we chose to incorporate in the system during the early periods of operation. These changes include:

- (1) Modifications to the carbide handling system by adding a Rietz Disintegrator which was installed to uniformly size the carbide lime slurry before it enters the system.
- (2) Modifications to the thickener circuit to improve settling characteristics of the calcium sulfite crystals by the addition of a flocculating system. The flocculents used in this system were Dow A-23 and Betz 1100. The best results were obtained with the latter.
- (3) Improvements to the top water washer in the demister system which was at the start of operations found to be inadequate.
- (4) Various piping and pH control changes.

Generally, the system has operated in a satisfactory manner and very few problems have been encountered to date. No major problems of scaling or plugging have been encountered in this 9 month period. Once we observed a minor build up of calcium sulfite scale that occurred when the scrubber was allowed inadvertently, in one instance, to run at a pH above the critical point of 6.2. This scale was easily removed by dissolving the crystals by lowering the overall pH of the system and no further build up or scaling has occurred. During these operations the SO₂ removal efficiencies averaged from 80% to 90%. The best operations were accomplished on a general pH input of approximately 8. This 8 pH area of control was selected to cover fluctuations in SO₂ concentration in the flue gas when the boiler operated on varying sulfur levels in the coal. The overall range of sulfur in the coal during these test periods were in the range of 1.5% to something greater than 4%. The particulate removal efficiency throughout these varying operating conditions fell well below the standard of .05 grains/Scf. Most of the data was in the range

of .02 and .035 grains, thus, producing a particulate efficiency well in excess of 99%. The inlet gas to the scrubber contained a particulate loading of about .2 to .4 grains.

The stoichiometric requirements of the system appears to approach 100% based on the total SO_2 removed or approximately 80 to 90 percent depending on the inlet SO_2 concentrations in the flue gas to the scrubber system.

As you are quite aware, this system has been operated in a closed loop mode and has never been out of that mode during any of our operations to date. This requirement of the system obviously only allows a small period of time to wash the demisters, since excessive use of wash water would upset the water balance and thus open the loop. Generally speaking, this washing operation at approximately 200 GPM of river water, is utilized 8 to 12 minutes every 8 hours. In the overall water balance, the only water that is removed from the system is that water which goes out with the reheated flue gas to protect the scrubber fan, which is in the circuit between the scrubber and the stack, and the water that is removed from the system as sludge from the filter circuit. The moisture content of the filter cake, which for our convenience is removed from the scrubber area by trucks, rather than some other method, approaches 55 to 65 percent water. This sludge is conveyed to a borrow pit approximately a mile from the plant where the sludge is mixed haphazardly with flyash to produce a rather solid landfill. In order that no one misunderstands, this is not in our opinion the best environmentally acceptable approach to a chemical fixation reaction. We feel that the sludge coming off either the drum filters or the higher percentage moisture content sludge from the thickener should be mixed with the dry flyash in the immediate vicinity of the plant where lime or other types of additives are mixed so that the resulting product can be pumped to the storage area. The water and the solids will then set-up in time to hopefully produce an environmentally acceptable solid landfill. By so doing, the water is chemically bonded in the new resulting solids thus protecting the ground water and eliminating the need to transport water back into the scrubber system. Much work remains in the deve-

lopement of an economically feasible and environmentally acceptable method of waste disposal.

When the unit is operating at full load the overall electrical load demand on the system is basically equivalent to approximately 2.3% of the output of the boiler. This is accomplished by a relatively low liquid to gas ratio in the entire system of 30 plus (15 to 18 per bed) and a pressure drop on the overall system of 14 to 15 inches. In future installations our efforts will be directed toward lower capital and operating costs.

Obviously, from the above discussion we have operated the system to date in a rather trouble free mode. No major problems have existed or have occurred which have been identified as general problem areas such as corrosion, erosion, scaling, plugging, and maintenance difficulties. For the most part, the demister banks from a visual observation appear as clean as they were when the system first went into operation last year.

The system as you probably are aware has a bypass feature which has worked extremely well under varying conditions. In fact, other than total bypassing of the system, it enables a proportioning of the gases for recycle as the boiler load varies from time to time. This provides extensive turn-down capabilities which gives the system almost unlimited control. To be sure the use of recycle increases power requirements on lower load conditions.

After the operations were temporarily suspended in late December 1973, a complete inspection of the system was made and very few areas of difficulties were observed. During this past summer, we used the system as the boiler requirements occurred. At no time during this period did any difficulties occur in the system. During this running period, we changed the operating mode to a degree in that a waste product, high in MgO , was inoculated into the carbide lime slurry to evaluate the effects on the system with increased levels of MgO . This of course, was done because MgO appears to have a greater effect on SO_2 removal rates than do the basic calcium hydroxides. The main reasons for these inoculations was to ascertain if higher

levels of SO_2 removals could be accomplished. The results were quite interesting. For the most part, the levels of SO_2 in the flue gas to the scrubber averaged approximately 2,000 ppm and the exit SO_2 concentrations maintained themselves well below 50 ppm. Generally, we had no trouble in operating in this mode.

Overall, from this paper, it is obvious that we have established some very encouraging results, however, it would be impossible for us to state that absolute reliability of the system has been established.

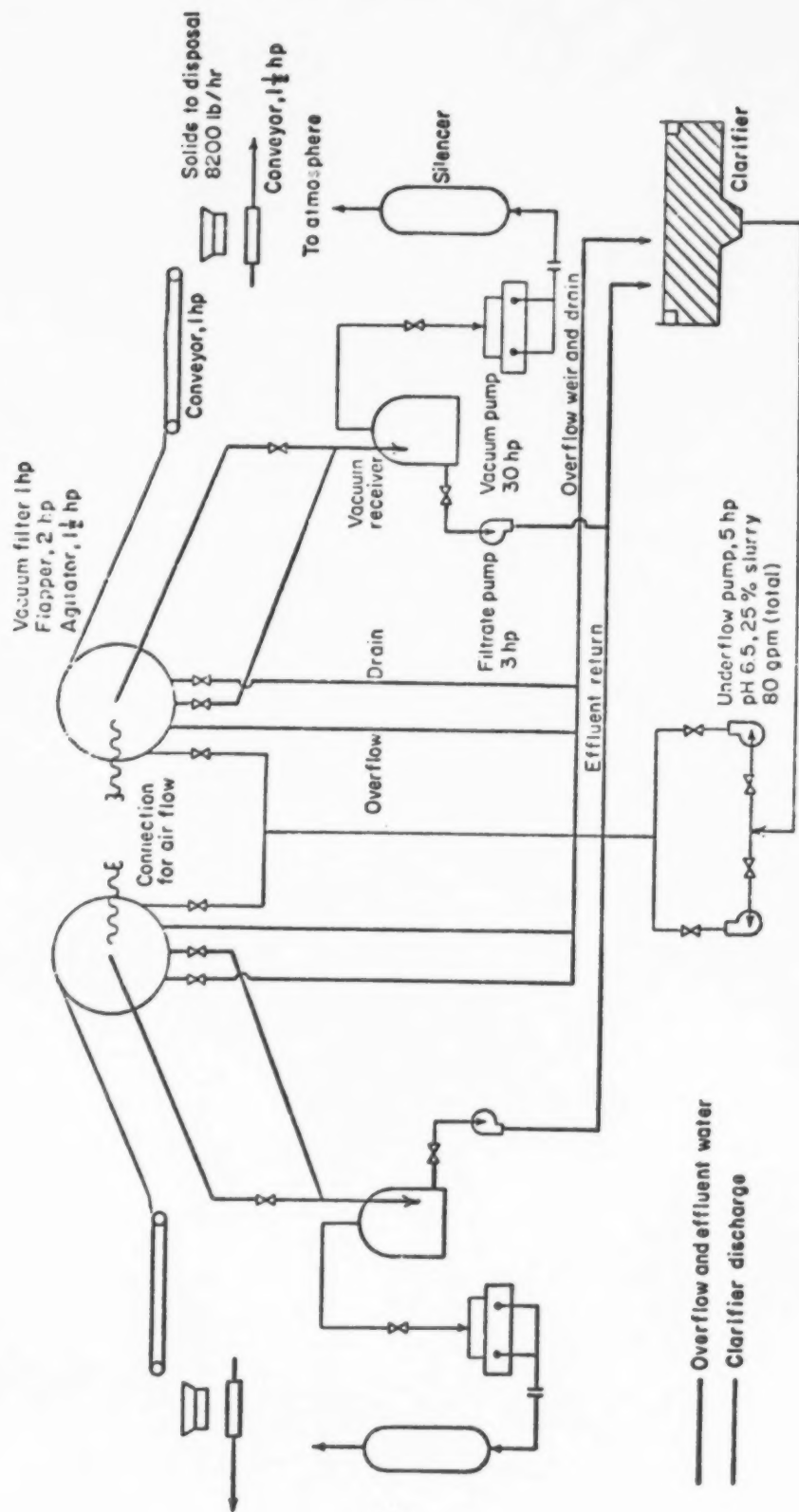
Now to other areas: Last year as required by the Clean Air Act of 1970, we submitted a complete compliance schedule for SO_2 removal systems for units requiring control at our plants known as Cane Run and Mill Creek. This plan basically proposed that the Primary Ambient Standard would be attained by July 1, 1977 and the Secondary Standard by July 1, 1979 or somewhat later. Obviously, this attempt did not satisfy the Clean Air Act since it exceeded the guidelines in time, even though, we felt it to be a very optimistic schedule.

After considerable time and many lengthy discussions with various regulatory bodies, the first phase of the plan was agreed upon and a variance was granted us by the local air pollution board since the time of completion extended beyond the limit of local regulations. The enclosed exhibit is a schedule of the installations included in the approved compliance plan. With this agreement in hand the next order of business was to request authority to expend funds by obtaining a "Certificate of Convenience and Necessity" from the Public Service Commission of Kentucky. Initially the KPSC had granted a certificate on a new source boiler of 425 MW at Mill Creek Unit #3 which is scheduled for operations in early 1977. The next three units in the plan were units which would be retrofits, known as Cane Run #4, #5 and #6. Initially, Cane Run #4 was requested followed by a request for Cane Run #5. In late 1973 a formal hearing was held on Cane Run #4 with a decision by the Commission on March 20 of this year. This order granted the Certificate of Convenience and Necessity for Cane Run #4 but delayed any further consideration on

other units until Cane Run #4 had been installed and had successfully operated for one year. Thus, the dilemma of non-compliance of the plan until some undetermined time frame beyond the sacred date of July 1, 1977.

In order to better clarify our position, we requested the KPSC for a rehearing of their order. At this rehearing in May and June, intervenors testified. They included the Environmental Protection Agency, the Edison Electric Institute, The Mayor of Louisville, the County Judge of Jefferson County, the Air Pollution Control District of Jefferson County, the Kentucky Department of Natural Resources and Environmental Protection, and Action for Clean Air (a local environmental group). As of this date, in late September, when this report is being written, we have had no ruling by the KPSC. However, in the meantime, as you can see from the approved compliance plan as noted in exhibit B we have exceeded some of the progress dates in the actual plan. In short, we have an enforcement order from the Air Pollution Control District of Jefferson County to proceed, but can not proceed until Public Service Commission approval is received.

EXHIBIT A



C-E AIR-QUALITY CONTROL SYSTEM
LOUISVILLE GAS & ELECTRIC COMPANY
PADDY'S RUN NO. 6 VACUUM-FILTER SCHEMATIC

EXHIBIT B

Louisville Gas and Electric Company
Appendix VI
Sulfur Dioxide Removal System Installation
Proposed Construction Schedule*

	<u>Order Equipment</u>	<u>Start Construction</u>	<u>Construction Completed</u>	<u>Start-up Date</u>	<u>Completion of Start-up Period & Compliance By</u>
Cane Run #4	5-20-74	9-20-74	3-1-76	4-1-76	6-1-76
Cane Run #5	5-20-74	12- 1-74	8-1-76	9-1-76	11-1-76
Cane Run #6	9- 1-74	2- 1-75	4-1-77	5-1-77	7-1-77
398 Mill Creek #3 (Approx.)	3-27-74	6- 1-75	4-1-77	5-1-77	7-1-77

The above schedule is predicated on manufacturers' current delivery estimates and does not consider potential time loss due to strikes, material shortages or other unforeseen developments beyond the control of the Company.

*Revised as of 3/21/74

DISPOSAL OF BY-PRODUCTS FROM
NON-REGENERABLE FLUE GAS
DESULFURIZATION SYSTEMS

A STATUS REPORT

by

J. Rossoff
R. C. Rossi
L. J. Bornstein
Environment and Urban Division
The Aerospace Corporation
El Segundo, California

and

J. W. Jones
Control Systems Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina

Presented at the
EPA Control Systems Laboratory
Symposium on Flue Gas Desulfurization
Atlanta, Georgia
November 4-7, 1974

This paper has been reviewed by the Environmental Protection Agency and approved for presentation. Approval does not signify that the contents necessarily reflect the views and policies of the Agency, nor does mention of trade names, commercial products, or commercial processes constitute endorsement or recommendation for use.

CONTENTS

	<u>Page</u>
1.0 Introduction	402
2.0 Summary	403
3.0 Potential Water Pollution Problems	406
4.0 Potential Land Reclamation Problems	414
5.0 Potential Disposal Problem Solutions	422
6.0 Disposal Cost Estimates	430
7.0 Shawnee Sludge Disposal Field Demonstration	441
8.0 References	442

DISPOSAL OF BY-PRODUCTS FROM
NON-REGENERABLE FLUE GAS DESULFURIZATION SYSTEMS

A STATUS REPORT

by

J. Rossoff
R. C. Rossi
L. J. Bornstein
Environment and Urban Division
The Aerospace Corporation
El Segundo, California

and

J. W. Jones
Control Systems Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina

ABSTRACT

A summary of an on-going Aerospace Corporation study for the EPA's Control Systems Laboratory to characterize flue gas desulfurization sludges and to assess methods for their environmentally sound disposal is given. The sludges studied were selected from non-regenerable scrubber systems using lime or limestone as the absorbent at several Eastern and Western coal-burning power plants. Chemical analyses have shown that untreated sludge liquors contained mercury, selenium, boron, chloride, sulfate, and total dissolved solids considerably in excess of water quality criteria (most of the constituents of concern, which without scrubbing are discharged in the fly ash and flue gas, originate in the coal). Since soil attenuation of selenium, boron, and chloride is known to be ineffective, and since the remaining sludge liquor constituents prevent direct discharge to streams, the need for environmental control of sludge disposal sites is strongly indicated. In addition, the water-retentive properties of raw (untreated) sludge present land reclamation difficulties. Currently available disposal methods include landfilling of sludge (which has been chemically treated for strength and the prevention of leaching of undesirable constituents to water supplies), and disposal of untreated sludge in ponds (lined with highly impermeable materials). Estimates from various sources indicate disposal costs of from \$2 to \$9/ton for chemical fixation and from \$2.50 to \$4.50/ton for ponding, excluding possible subsequent pond reclamation costs. An average disposal cost of \$5/ton is equated to 1.1 mills/kwhr. Other less developed disposal techniques include drying, and conversion to gypsum.

DISPOSAL OF BY-PRODUCTS FROM NON-REGENERABLE FLUE GAS DESULFURIZATION SYSTEMS - A STATUS REPORT

1.0 INTRODUCTION

The Aerospace Corporation sludge disposal study discussed in this paper is in its second year of a two-year program which began in November 1972 under EPA Contract 68-02-1010. The results of the first year's work is described in detail in an initial report¹. This paper up-dates that work in a summary form.

Sludge samples have been obtained from the EPA/TVA Shawnee limestone and lime prototype scrubbers, the Southern California Edison Mohave Station limestone pilot scale scrubber, and the General Motors Parma facility double alkali scrubber. The major portion of the analyses of all these samples has been completed except for the double alkali samples. Additionally, technical and economic assessments are being made of power plant and commercial processor operations being conducted to develop environmentally sound sludge disposal techniques. Chemically conditioned sludge samples from some of these sources, namely, Chemfix, IU Conversion Systems, and Duquesne Light Company which uses a Dravo Corporation additive have been received and are being analyzed. Reviews are being made continually of state and federal standards which govern water quality and the potential land disposal of sludges. Present plans by the EPA are to expand the Aerospace study to include the sampling of sludges from four other scrubber systems to increase the data base being established in this study.

A sludge disposal field demonstration program of untreated and treated sludges is being initiated by the EPA at the TVA Shawnee site in Paducah, Kentucky. Analyses of sludges, leachate, and ground water from that site will be conducted during the next two years in the Aerospace program and will be integrated with the Aerospace study data and other data as appropriate.

2.0 SUMMARY

In this study, sludge samples have been analyzed from three scrubbers involving lime/Eastern coal, limestone/Eastern coal and limestone/Western coal. Additionally, two sludges from other scrubbers (limestone/Western coal and lime/Eastern coal) were analyzed in another study for the U.S. EPA National Research Center, Corvallis, Oregon (EPA Research Grant R802853-01-1) and assessed as a part of the total program. Chemical characterizations of the sludge liquors were performed to identify the concentrations of nine trace metals of interest and seven major soluble species, as well as the pH and total dissolved solids (TDS) for each of the samples collected. (Input materials such as coal, fly ash, make-up water, lime and limestone were also characterized to identify source of constituents found in the sludges). A preliminary assessment was made of these samples with regard to any potential environmental problem that might be posed if these sludge liquors entered water supplies. This assessment was made by comparing the chemical characterizations with the criteria established by the states, the U.S. Public Health Services Drinking Water Standards - 1962, and the EPA Proposed Public Water Supply Intake Criteria - October 1973. The latter criteria were used in the final analysis at this stage of the study.

In assessing the potential impact of trace metals, it was found that in each of the sludge liquors at least one of the following trace metals exceeded the EPA proposed criteria: arsenic, cadmium, chromium, lead, mercury, and selenium. Except for mercury and selenium, these trace metals exceeded the criteria by not more than a factor of 5. Mercury and selenium exceeded the criteria in each of the sludges analyzed by more than an order of magnitude. Comparing the concentrations of the major soluble species with the criteria, it was found that particular excesses exist for chloride, sulfate and TDS for each of the samples analyzed. To date, only one sample has been analyzed for boron, which was also found to be in excess. It should be noted that most of the constituents of concern, which without scrubbing are discharged in the fly ash and flue gas, originate in the coal.

From the assessment just described, it appears that in all of the power plant sludge liquors analyzed, water quality criteria are appreciably exceeded for mercury, selenium, boron, chloride, sulfate and TDS. Attenuation of chemical species by soil is widely accepted in many land

disposal practices; however, attenuation of three of the species discussed above, namely, selenium, boron and chloride is known to be ineffective. Therefore, the concentrations of these three species strongly indicate that environmentally sound techniques such as lined ponds or chemical fixation are needed for scrubber sludge disposal.

Detailed physical characterizations of various samples have also been performed under the Aerospace Corporation study to determine properties such as: ease of dewatering/settling, bulk density, viscosity, load bearing strength and permeability. Some significant results relating to land-use capabilities are as follows: scrubber sludges are highly water-retentive and have little or no bearing strength as produced; they are thixotropic; they can be dewatered and/or compacted to produce appreciable bearing strength and low coefficients of permeability, but after being dewatered, they easily rewet in the presence of water.

A sulfite* sludge can be dewatered in a thickener to between 35 and 45 percent solids and will not increase its solids content significantly by settling; however, when allowed to drain freely, its solids content will increase to about 50 percent. A filter or centrifuge will produce a solids content of about 50-55 percent. In contrast a sulfate* sludge which can be dewatered to 35-45 percent solids in a thickener, will increase its solids content to about 65 percent upon settling, and when freely drained or centrifuged, will dewater to about 75 percent solids. It can be dewatered to about 80 percent solids by filtration.

A better comparison of physical properties can be made when sulfite and sulfate sludges containing equivalent moisture content are compared. For example, at 65 percent solids content, sludges will support personnel, and at solids content greater than 70 percent, sludges will support heavy equipment. A settled sulfate sludge can be easily compacted such that it will readily support personnel and equipment, but sulfite sludges by comparison are much more difficult to dewater to the required degree of dryness for compaction.

From the preceding discussion, it appears that raw sludges or raw sludge ponds which have been dewatered could be used for structural purposes.

* Sulfite and sulfate sludges described herein apply specifically to TVA Shawnee and S.C. Edison Mohave pilot plant sludges defined in Figures 1, 2 and 3.

However, the cost of dewatering the sludge and maintaining the site at the necessary degree of dryness is strongly dependent on the raw sludge properties, and may or may not compare favorably with the cost of an alternative disposal technique such as chemical fixation. Data from chemical fixation processors indicate that treated sludge quickly attains a compressive strength above the limit for hard clay or compacted fine sand as defined in the Uniform Building Code. This conditioned sludge would have the capability to make a disposal site reclaimable for either structural or recreational use.

Based on the sludge properties just described, the best practical methods currently available for sludge disposal are ponding, and chemical fixation for landfilling. Ponding which may be environmentally sound with regard to water quality consists of lining the containment basin with an impervious material such as clay, plastic or rubber to prevent the seepage of sludge liquors to water supplies. There are certain problems attendant to this approach which are still being considered; these include the inability of the liners to last indefinitely and the difficulty of reclaiming the land after the site is abandoned. A more technically sound approach is that offered by industry in which the material is chemically conditioned so that leaching of undesirable constituents is minimized and the material can be easily used to produce a structural landfill.

Preliminary cost estimates of these disposal alternatives, assuming a sludge consisting of 50 percent solids, place the cost of ponding in the range of \$2.50 to \$4.50 per ton.* Some additional cost to provide permanent environmental protection or land reclamation if desired may be incurred at the end of the service life of the pond. The cost of disposal by fixation currently is quoted in the range of \$2.00 to \$6.00 per ton by the fixation processors and approximately \$7.50 to \$9.00 per ton by power companies performing their own disposal task. The cost of ponding appears to be about half that of fixation; however, ponding is not necessarily a permanent disposal technique.

A disposal cost of \$2.50 per ton, in terms of power produced, is equivalent to 0.56 mills/kwhr; \$5.00 per ton is equivalent to 1.12 mills/kwhr. This example is based on the use of a limestone scrubber, 85 percent

* All costs quoted herein are on a wet sludge basis.

SO₂ removal, 1.2 CaCO₃/SO₂ mole ratio, 50 percent solids sludge plus ash at a station burning coal that has 3 percent sulfur and 12 percent ash, at 0.88 lb/kwhr.

3.0 POTENTIAL WATER POLLUTION PROBLEMS

3.1 WATER QUALITY

The disposal of scrubber sludges presents potential water pollution and land reclamation problems because of soluble species and dewatering difficulties. The measure of those problems is defined by authoritative water quality standards and solid waste disposal regulations, or by judgment of regulatory officials when standards are nonapplicable or nonexistent. The most nearly applicable water quality criteria that exist for determining the environmental acceptability of sludge disposal are given in Table 1. This table shows the water quality criteria of the U.S. Public Health Service (USPHS) - 1962 Drinking Water Standards², the 1973 State of Illinois criteria (an example of one of the more comprehensive of all state criteria) for both public and food processing water supply (PWS) and water effluent discharge standards³, and the 1973 EPA Proposed Criteria for Water Quality⁴.

Regarding the quality of surface waters, it should be noted that all states, in conformance with Federal EPA guidelines, do not consider dilution of an effluent as an acceptable method of treating wastes in order to meet the effluent standards. Regarding the quality of underground waters, in general, states which do not have specific regulations concerning the quality of underground water apply stream or drinking water standards to underground water for lack of more definitive regulations. One state which does regulate underground waters is Illinois, which requires that underground waters that are a present or potential source of water for public or food processing supply shall also meet the standards for those supplies.

Standards for the regulation of solid waste disposal, as related to water quality, are much less specific than water quality criteria. Regulations ordinarily limit the disposal of toxic, hazardous, and harmful substances by requiring that those materials not be disposed of on the ground, or underground, without the approval of a designated authority such as the State Department of Health, the State Health Commissioner, County Health Commissioner, or the State Department of Commerce. The

TABLE 1. WATER QUALITY CRITERIA

	CONCENTRATION, ppm			
	USPHS - 1962 DRINKING WATER STANDARDS	ILLINOIS - 1973		EPA PROPOSED PUBLIC WATER SUPPLY INTAKE OCT 1973
		PUBLIC AND FOOD ^a PROCESSING WATER SUPPLY	EFFLUENT ^b DISCHARGE STANDARDS	
ARSENIC	0.05	0.01	0.25	0.1
BORON	--	1.0	--	1.0
CADMIUM	0.01	0.01	0.15	0.01
CHROMIUM (total)	--	--	--	0.05
COPPER	1.0	0.02	1.0	1.0
LEAD	0.05	0.05	0.1	0.05
MERCURY	--	0.0005	0.0005	0.002
SELENIUM	0.01	0.01	1.0	0.01
ZINC	5	1.0	1.0	5.0
CHLORIDE	250	250	--	250
FLUORIDE	2.4 AT 50° F 1.4 AT 90° F	1.4	2.5	--
SULFATE	250	250	--	250
TDS	500	500	3500 ^c	NO LIMIT ^d
pH	--	6.5 - 9.0	5 - 10	5.0 - 9.0

^aPoint at which water is withdrawn for treatment and distribution

^bBased upon 24-hour composite sample. No contaminant shall at any time exceed five (5) times the concentration limit

^cNot more than 750 ppm above background concentration limits unless caused by pollution abatement practices

^d"No limit" indicates that insufficient data existed for prescribing limits

survey undertaken to date indicates that those offices, when judging the disposal of sludges, base their judgment on sludge quality as related to the local or state water quality criteria.

The limitations imposed on sludge disposal have been further intensified by the recent passage of legislation and the potential passage of others. For example, the Federal Water Pollution Control Act Amendments of 1972⁵ essentially require the revision and re-approval of all state water quality standards: to change applicability from interstate waters to all waters, to prohibit or limit direct discharges of pollutants to streams, to limit subsurface disposal, and to include Federal EPA-prescribed criteria for specific toxic elements. Additionally, the federal drinking water standards are being revised, and legislation is in process to develop new limitations for the land disposal of solid wastes.

3.2 SLUDGE SAMPLING

Sludge samples analyzed to date were taken from three scrubbers at two different power plants representing both Eastern and Western coals and the lime and limestone processes. These are: EPA/TVA Shawnee TCA scrubber with limestone, EPA/TVA Shawnee Venturi scrubber with lime, and Southern California Edison (SCE) Mohave TCA pilot scrubber with limestone. Additionally, sludge liquors from two other plants (an Eastern utility plant lime scrubber, and a Western utility plant limestone scrubber) have been characterized in a separate study for the EPA NRC, Corvallis, Oregon, and assessed along with the data determined in this program. Samples were taken at various points within the scrubbing system so that the fate of trace elements and major species of interest could be "observed" as they passed through the system. In addition, samples of process ingredients -- coal, process water, fly ash, lime, and limestone -- were taken to identify the source of the elements entering the system. Samples were taken over a short period of time in an attempt to minimize the effect of system fluctuations.

3.3 CHEMICAL CHARACTERIZATION

Chemical analysis of the sludge samples consisted of the determination of:

a) magnesium and trace metals by atomic absorption spectrophotometry, b) calcium by wet chemical methods, c) sulfate by turbidimetry, and d) chloride, fluoride, and sulfite by specific ion electrode. Results of these analyses are presented in Tables 2 and 3 for the various power plant sludges analyzed. In each case (except for Plant A), the analysis is for the liquor fraction of the sludge representing the condition in which the sludge is likely to be disposed, e.g., clarifier underflow, filter cake or centrifuge cake (Plant A liquor is scrubber effluent and does not represent actual disposal material).

Since these analyses represent sludge liquors from a variety of scrubber systems, certain qualifications must be made for accurate interpretations of the data. Four principal system conditions influence the concentrations of soluble species in system liquors. The first is the "tightness" of the scrubber system liquor loop. The tighter the system, the less liquor is lost with the discharge from the recirculation system to the disposal area. Consequently, less make-up water is required and less dilution is possible. An example of the effects of this condition can be seen in Table 3 where it is shown that the concentrations of chlorides and TDS in the SCE Mohave liquor are at least an order of magnitude greater than those in the Plant A liquor. The Mohave pilot scrubber had a tightly closed loop system, whereas the Plant A scrubber was running open loop and consequently experienced a high degree of liquor dilution.

If it were theoretically possible to dewater the sludge solids completely and return all liquor to the system, no make-up water would be required for water balance except to make-up for evaporation. Soluble species would build up in concentration until they reach a saturation level resulting in precipitation and subsequent removal with the solids. From the point of view of minimizing the volume of waste material, an absolute separation of solid and liquid might appear desirable, but from the point of view of a reliable system operation, an excessively tight liquor loop could present major difficulties in the operation of the system. In practice, disposal of liquor with solids is not only necessary but unavoidable.

The second principal system condition that can affect the concentration of soluble species is related to the operational differences in scrubbing flue

TABLE 2. TRACE METALS IN SLUDGE LIQUORS^a

(concentration in ppm^b)

SCRUBBER	TVA SHAWNEE LIME	TVA SHAWNEE LIMESTONE	SCE MOHAVE LIMESTONE	PLANT A ^c LIMESTONE (western coal) >100	PLANT B LIME (eastern coal) >100	WATER QUALITY CRITERIA EPA PROPOSED PUBLIC WATER SUPPLY INTAKE OCT 1973
• SIZE, Mw	10	10	<1	15	35-40 ^d	
• PERCENT SOLIDS IN DISCHARGE	50-55	35-40	75-80			
ARSENIC	0.02	0.20	0.03	<0.004	0.085	0.1
BERYLLIUM	<0.002	0.01	<0.001	0.18	0.012	--
CADMIUM	0.10	0.005	0.05	0.01	0.023	0.01
CHROMIUM (total)	0.03	0.15	<0.005	0.21	0.040	0.05
COPPER	<0.002	0.02	0.08	0.20	0.048	1.0
LEAD	0.05	0.1	0.01	0.02	0.18	0.05
MERCURY	<0.001	0.06	0.0012	0.12	0.045	0.002
SELENIUM	<0.02	0.30	0.12	2.5	0.80	0.01
ZINC	0.08	0.30	0.12	0.12	0.09	5.0
pH	9	8	7	4	9	5 TO 9

^a All sludge liquor samples, except those from SCE Mohave, contain significant amounts of fly ash

^b Equivalent to mg/l

^c Plant A sample is scrubber effluent and does not represent actual disposal material

^d An underdetermined amount of clarifier overflow liquor is discharged from this system, making the effective percent solids in discharge somewhat lower than value shown

TABLE 3. MAJOR SPECIES IN SLUDGE LIQUORS^a

SCRUBBER	(concentration in ppm ^b)					WATER QUALITY CRITERIA EPA PROPOSED PUBLIC WATER SUPPLY INTAKE OCT 1973
	TVA SHAWNEE LIME	TVA SHAWNEE LIMESTONE	SCE MOHAVE LIMESTONE	PLANT A ^c LIMESTONE (western coal) > 100	PLANT B LIME (eastern coal) > 100	
• SIZE, Mw	10	10	<1	15	35-40 ^d	
• PERCENT SOLIDS IN DISCHARGE	50-55	35-40	75-80			
CALCIUM	2520	1600	1400	700	1400	--
MAGNESIUM	25	600	-- ^e	δ	410	--
BORON	40.8	-- ^e	-- ^e	-- ^e	-- ^e	1.0
CHLORIDE	5000	2500	30,000	1400	2700	250
FLUORIDE	3.3	3.4	3.1	0.6	2.6	--
SULFATE	800	2000	2500	1150	2250	250
SULFITE	0.9	110	<0.1	<0.7	20	--
TDS ^f	9000	7000	70,000	3500	7000	NO LIMIT ^g

^aAll sludge liquor samples, except those from SCE Mohave, contain significant amounts of fly ash
^bEquivalent to mg/l

^cPlant A sample is scrubber effluent and does not represent actual disposal material
^dAn undetermined amount of clarifier overflow liquor is discharged from this system, making the effective percent solids in discharge somewhat lower than value shown

^eTo be determined

^fIncludes other soluble species (e.g., sodium and potassium)

^g"No Limit" indicates that insufficient data existed for prescribing limits

gas from Eastern and Western coals. In all cases, the solids content of the waste bleed stream from the scrubber loop is typically 10 percent (solids contents from 8 to 15 percent have been measured and appear to depend on the presence of fly ash in the liquor and system design variables associated with relative liquor holding capacity). However, for equivalent system operation and equivalent sulfur scrubbing efficiency, the sludge production rate (and therefore the sludge bleed rate, excluding ash) from the scrubber recirculation loop for low sulfur Western coal is considerably less than that for a high sulfur Eastern coal. Therefore, less occluded liquor would be bled from the Western system than the Eastern system. Since absorption of soluble species takes place with each liquor pass through the scrubber, and since the scrubber liquor "residence time" is greater in the Western coal system, the concentration of soluble species is greater. The ratio of the lowest Western coal sulfur content to the highest Eastern coal concentrations analyzed and reported here is 1:8 which implies that the scrubber loop liquor residence time is about 8 times longer in this low sulfur system (see TDS for Mohave pilot and Shawnee liquors in Table 3). This is an extreme condition which compares an Eastern high sulfur coal sludge which includes fly ash, with a Western coal sludge that contains an insignificant amount of fly ash. If fly ash had been collected along with SO_2 in the Mohave scrubber, this effect of sludge liquor extended residence time would not have been as pronounced.

The third principal system condition affecting the concentration of soluble species pertains to the system design for collection of fly ash. Most of the trace metals in the sludge originate from fly ash simultaneously scrubbed from the flue gas; therefore, the scrubber systems that do not collect fly ash should be expected to have significantly lower concentrations of trace metals than those systems that do collect fly ash.

The fourth principal system condition affecting the concentration of soluble species is the pH at which the system operates. For example, the lime scrubbing process operates at a higher bleed pH than the limestone process.

As a consequence, the soluble element content in lime sludge liquors appears to be consistently lower than the same element content in limestone scrubbing systems. This trend can be most easily seen when comparing the lime and limestone systems at Shawnee where input flue gases are generated from the same boiler (see Table 2). In each case, the concentration of trace elements in the lime system is less than that in the limestone system. These data suggest that the more environmentally acceptable sludges (from the point of view of soluble trace metals in sludge liquors) are those disposed of at the high pH more typical of lime scrubbing systems.

Preliminary correlation between the soluble element content of the sludges and their concentrations in coal has been made. Arsenic, lead, and mercury are found in higher concentrations in Eastern coals than in Western coals, and their concentration in Eastern power plant sludges generally is also higher. Similarly, Western coals have higher concentrations of copper and selenium and these elements are generally found in greater concentrations in Western power plant sludges.

3.4 ASSESSMENT

The environmental acceptability of sulfur sludge liquors can be assessed by the comparison of the analyses determined on actual sludges presented in Tables 2 and 3 with the water quality criteria of various regulatory agencies presented in Table 1. For the trace metals (see Table 2), all elements analyzed except copper and zinc exceed water quality criteria in at least one power plant. The remaining trace elements, except mercury and selenium, exceed the criteria by not more than a factor of five, with one exception (cadmium - Shawnee lime). Mercury and selenium exceed these criteria for each of the power plants, and the excess is greater than an order of magnitude.

Comparing Tables 1 and 3 for major soluble species, it is seen that water quality criteria are surpassed in several cases. Particular excesses exist for chloride, sulfate and total dissolved solids for each of the power plants. Boron is in excess based on the one sample analyzed so far. The

values for sulfate ions for any given plant will not vary too widely from those given here because they are controlled by the chemistry of the scrubbing system. On the other hand, the chloride ion and total dissolved solids values can vary by an order of magnitude or more and are dependent on factors such as "tightness" of the loop, coal composition, and sludge bleed rate.

Summarizing the above comparisons of trace metals and soluble species against water quality criteria, in all of sludge liquors analyzed, water quality criteria were appreciably exceeded for mercury, selenium, boron, chloride, sulfate and TDS. It should be noted that most of the constituents of concern, which without scrubbing are discharged in the fly ash and flue gas, originate in the coal. Attenuation of chemical species by soils is widely accepted in many disposal practices. However, soil attenuation of selenium, boron, and chloride is known to be ineffective. Therefore, even if excessive concentrations of other constituents were acceptable because of soil attenuation effects, the presence of these three species in concentrations in excess of water quality criteria strongly indicates that environmentally sound techniques need to be employed for scrubber sludge disposal.

4.0 POTENTIAL LAND RECLAMATION PROBLEMS

4.1 LAND USE CRITERIA

Criteria identifying the load support limits to be attained on top of a filled sludge pond for eventual use as structural or recreational land are not included in either the state solid waste disposal regulations or building code requirements. However, the Uniform Building Code⁶ shows that the range of allowable soil pressure for building construction is from 2441 kg/sq m (500 psf) to 39,064 kg/sq m (8000 psf) depending on the type of soil and the depth of footing used. Samples of the weakest and strongest soils considered for structural usage are loose inorganic sand silt mixtures and hard clay, respectively. Materials like loose organic sand and silt mixtures and muck or bay mud are considered to have "zero" allowable soil pressure, but it is accepted practice to reclaim municipal landfills (e.g., organic soils) for recreational land use.

4.2 PHYSICAL CHARACTERISTICS

The physical characteristics of power plant sludges are primarily dependent on the percent moisture. However, the degree to which a sludge can be dewatered is dependent on the relative quantities of the major phases present, i. e., calcium sulfite, calcium sulfate, calcium carbonate and fly ash. Among these, the dewatering capabilities of a sludge depend mostly on the degree of oxidation of the sulfite phase. Primarily as a consequence of the platelet-like nature of the sulfite crystals, water becomes entrapped between the crystalline plates and is not easily or readily removed by physical dewatering techniques, such as filtration or centrifugation. In contrast, the sulfate crystals are "blocky" and do not trap water. A sulfite sludge can be dewatered in a thickener to 35-45 percent solids and will not increase solids content significantly by settling. A filter or centrifuge will increase solids content to about 50-55 percent. A sulfate sludge dewatered to 35-45 percent solids will increase its solids content to 60-65 percent upon settling, 75 percent when freely drained or centrifuged, and about 80 percent by filtration. The presence of significant amounts of fly ash will slightly alter the dewatering characteristics by increasing the solids content of sulfite sludges but will decrease slightly the solids content of sulfate sludges. Since some degree of oxidation exists in all power plant sludges, observed behavior tends to fall between these values depending on the relative quantities of the phases present. Figures 1 and 2 show the dewatering behavior of two specific sludges, Shawnee limestone sludge and SCE Mohave limestone sludge.

Since all physical properties of a sludge are dependent on moisture content, it is not possible to relate directly the behavior of different sludges at similar stages. For example, a filtered sulfate sludge will produce a cake that is firm, easily compacted and can readily support personnel and equipment. In contrast, a filtered sulfite sludge will produce a cake that may have a solid-like appearance but has a consistency much like toothpaste. A better comparison of physical properties can be made when sludges containing equivalent moisture content are compared. For example, at 65 percent solids content, sludges are firm enough to support personnel,

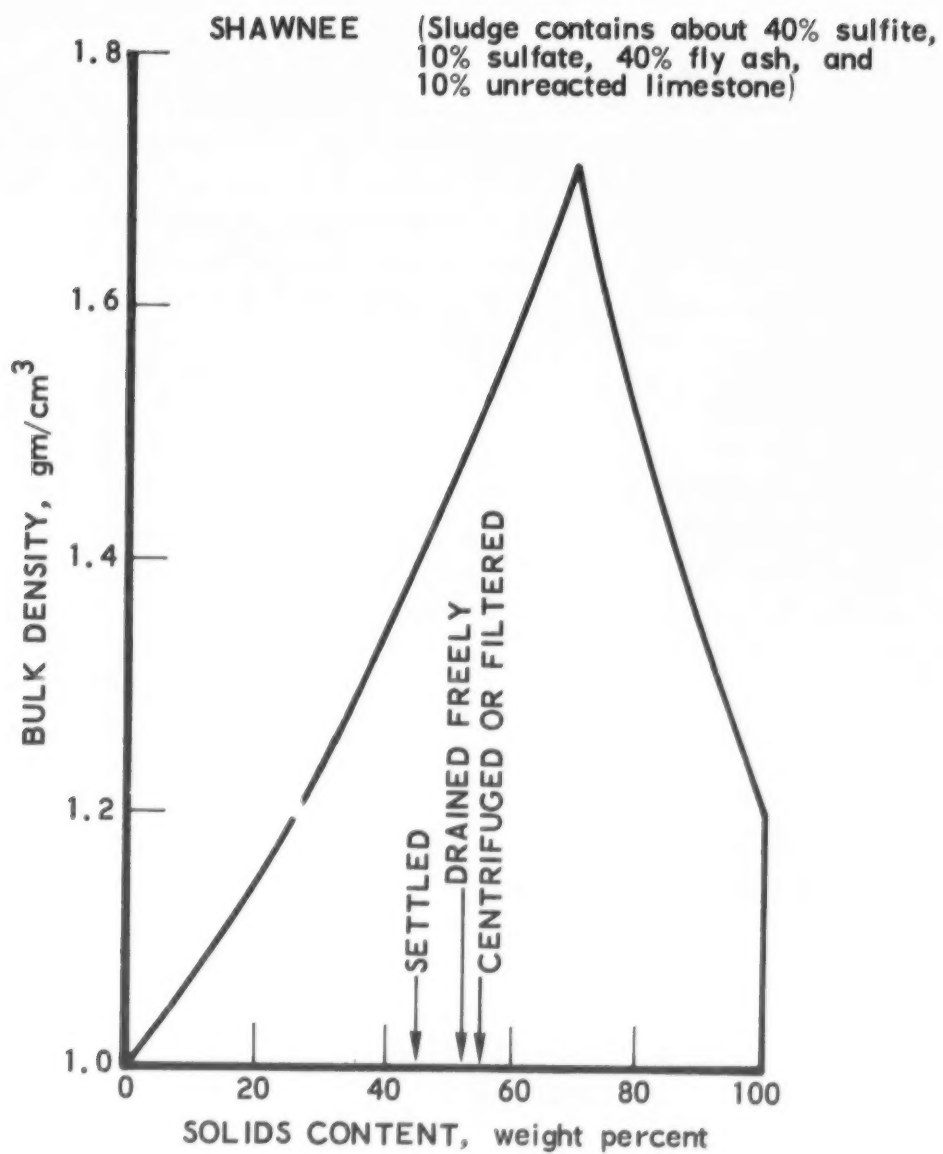


Figure 1. EPA/TVA Shawnee limestone sludge bulk densities

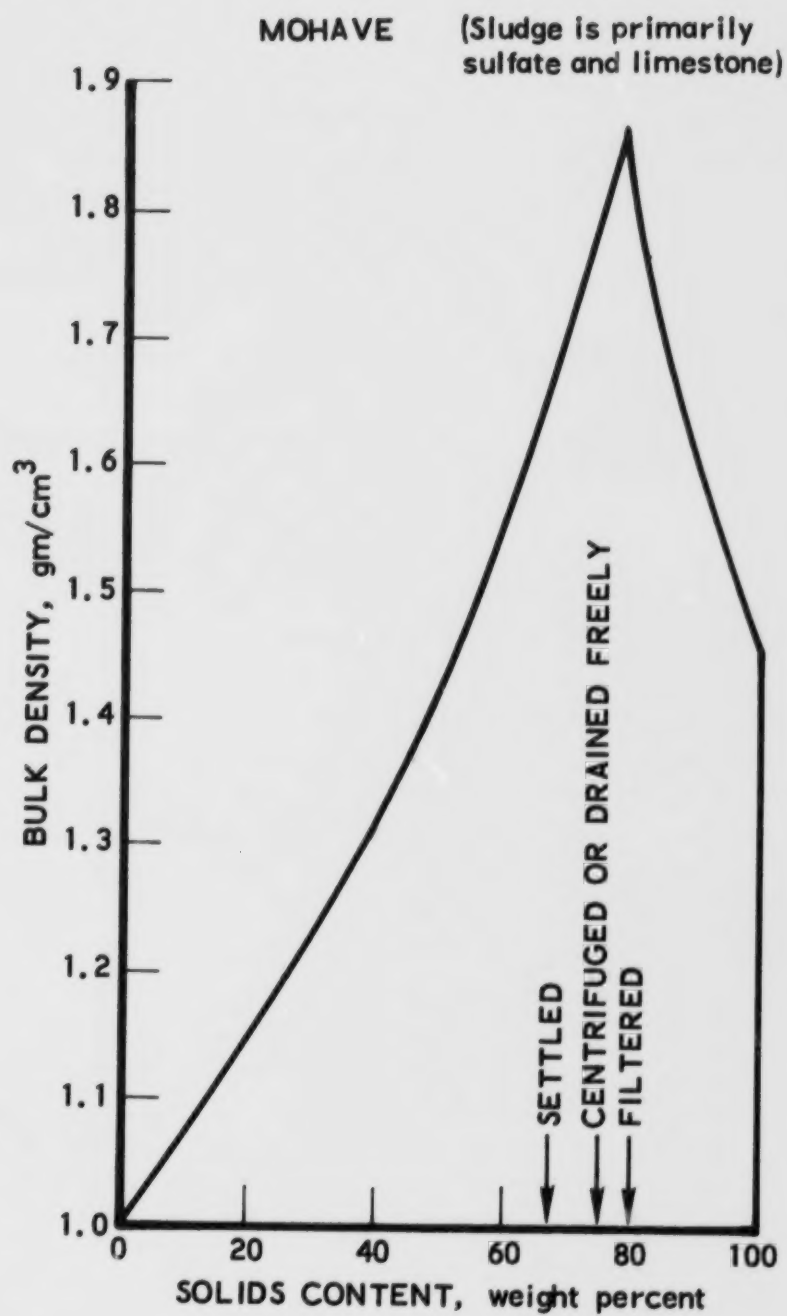


Figure 2. SCE Mohave limestone sludge bulk densities

and at solids content greater than 70 percent, equipment can be supported. The thixotropic behavior of the sludge is a direct function of sulfite content. Hence, while sludges with equivalent moisture content are capable of supporting equivalent loads, the work applied to a sulfite sludge in walking, or moving equipment, will tend to liquefy the sludge whereas a sulfate sludge will tend to remain firm. The presence of fly ash in the sulfite sludge will tend to firm it somewhat, but has no observable effect on a sulfate sludge. Again an example is made of the TVA Shawnee limestone and SCE Mohave limestone sludges (Figure 3) where the load carrying capacity of the sludges are compared at equivalent moisture content.

When a sludge is dewatered or dried to a point at which air voids are created, the sludge can be compacted so as to increase its bulk density. The point at which this behavior occurs varies among sulfite and sulfate sludges. Because of the particle agglomeration nature of sulfite crystals, sulfite sludges will create air voids at a solids contents as low as 60-65 percent. Sulfate sludges develop air voids at a solids content of about 80 percent. The presence of fly ash in the sludge will affect the sulfite sludge by increasing the solids concentration at which air void formation is reached. As in other properties just mentioned, fly ash has little effect on sulfate sludges. The amount of compaction is dependent upon many factors depending on phase compositions, moisture content, and compaction force, but dry bulk densities up to 1.5 gm/cm^3 (93 lb/ft^3) can be obtained in many sludges. This represents a 25 percent volume reduction on a dry solids basis for sulfite sludges but only a 7 percent reduction for sulfate sludges. Although the efficiency of field compaction may not be as high as that in the laboratory, relative volume reductions by compaction between the two extreme types of sludges will be generally the same. Figure 4 shows the compaction capabilities for typical sulfite and sulfate sludges as a function of moisture content. The moisture content for optimum compaction is about 10 percent for sulfate sludges but about 20 percent for sulfite sludges. This greater moisture requirement for sulfites is primarily a consequence of the entrapment of moisture between the platelet-like crystals that make up particle agglomerates.

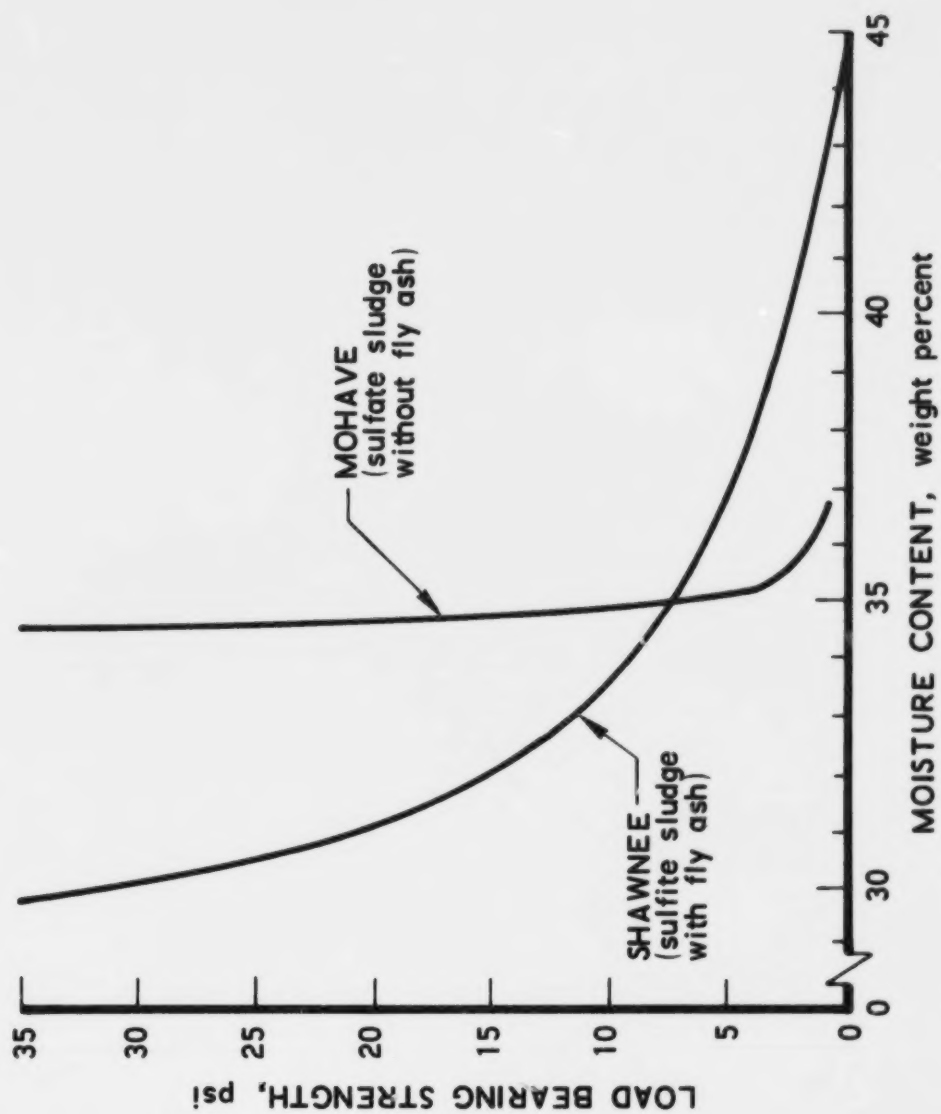


Figure 3. Sludge compaction strength

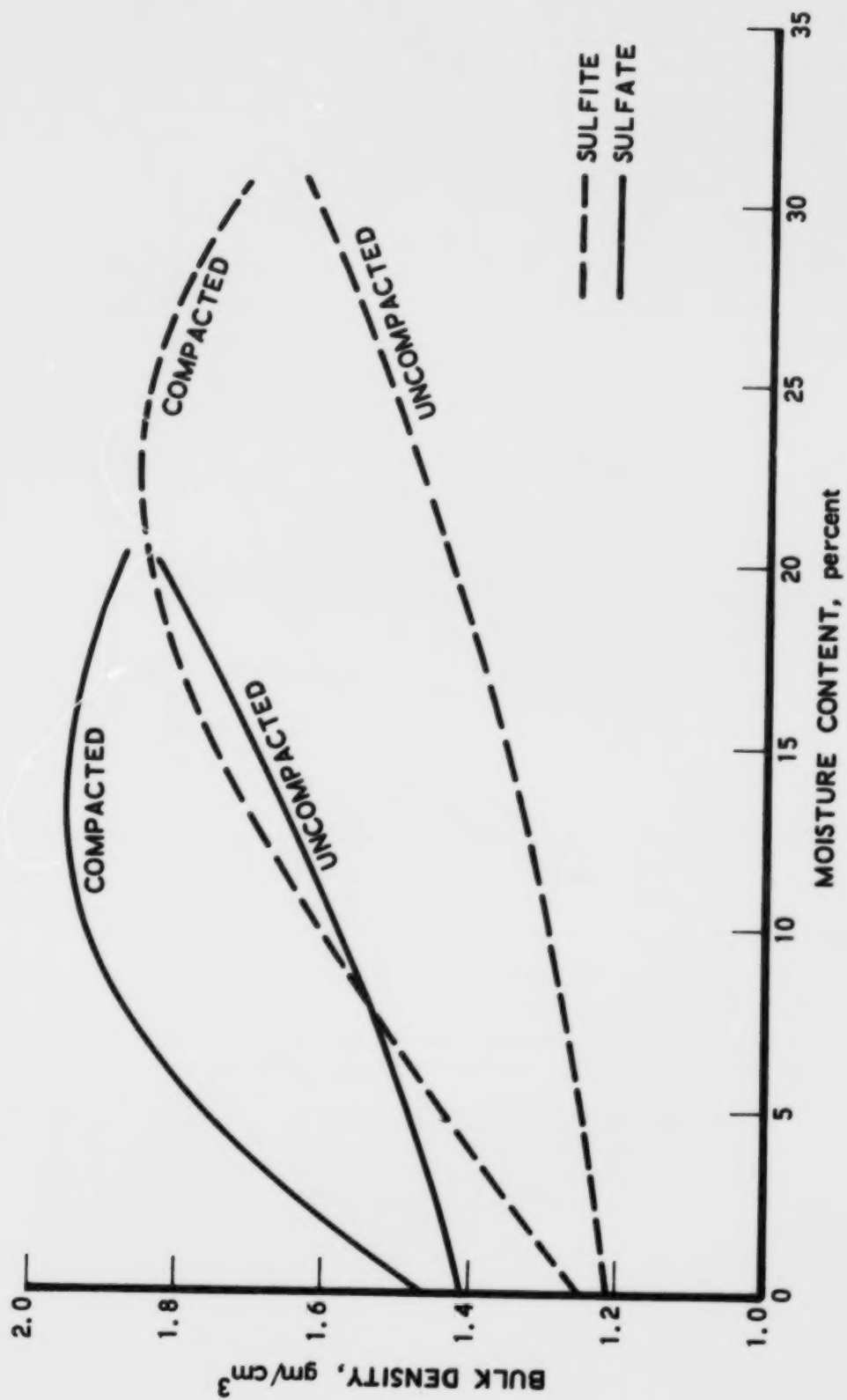


Figure 4. Preliminary compaction capabilities of sulfur sludges

The density of sludge is of primary importance because it is the major determinant of its compaction strength, compressibility, and permeability; all these factors affect the environmental acceptance of the sludge for disposal. The density also determines the quantity of sludge that can be stored in a given disposal area, thus affecting disposal costs. However, cost savings gained by higher density filling can be offset by the additional cost of compaction. On the other hand, if high density fill would render the disposal area to a greater economic use after reclamation, this factor must also be considered in choosing disposal techniques.

From the dewatering data discussed earlier, it can be seen that a sulfate sludge dewatered by filtration or centrifugation will have a moisture content very near the requirement for optimum compaction, although the volume gain by compaction is minimal. Sulfite sludges do not dewater as well as sulfate sludges and will require subsequent dewatering such as combining with dry additives, or thermal drying to attain the moisture content for optimum compaction. It should be noted that although a sludge can be dewatered to a degree such that it will support personnel and equipment, it will readily rewet in the presence of water. The gain of moisture in rewetting is limited by the volume of the voids in the sludge. Therefore, a rewetted compacted sludge will not return to its original moisture content or volume. By contrast, a rewetted uncompacted sludge will have properties nearly identical to those it had in its original wet state. Studies are continuing to determine the degree to which different rewetted compacted sludges approach their original state.

An additional physical parameter affecting disposal technology is the permeability of the sludge to leachate waters. Permeability among the sludges can vary over several orders of magnitude depending on phase composition and degree of compaction. Sludge that is allowed to settle and drain freely will have a permeability value equivalent to that of fly ash, about 10^{-4} cm/sec. Sulfate sludge in a settled state will drain more readily than sulfite sludge in the same state. When these sludges are compacted, drainability is reduced. The addition of fly ash also reduces the drainability of sludges. The drainability of sulfite sludges is reduced

only slightly, but the drainability of sulfate sludges is reduced by nearly an order of magnitude. A permeability value as low as 10^{-6} cm/sec has been obtained for a compacted sulfate sludge containing fly ash. This appears to be the minimum permeability obtainable for untreated sludge.

4.3 ASSESSMENT

Laboratory tests have shown that raw sludge which has been dewatered to about 65-70 percent solids can support loads in the range of 4883 kg/sq m (1000 psf) to 24,415 kg/sq m (5000 psf). Thus, it appears that raw sludge or raw sludge ponds which have been dewatered could be used for structural purposes per the Uniform Building Code. However, the cost of dewatering the sludge and maintaining the site at the necessary degree of dryness is strongly dependent on the raw sludge properties, and may or may not compare favorably with the cost of an alternative disposal technique such as chemical fixation.

Data from chemical fixation processors indicate that treated sludge quickly attains a compressive strength above the maximum limit for compacted fine sand or hard clay; i.e., 39,064 kg/sq m (8000 psf). This material has the capability to make a disposal site reclaimable for either structural or recreational use.

5.0 POTENTIAL DISPOSAL PROBLEM SOLUTIONS

5.1 DISPOSAL ALTERNATIVES

The choice of disposal alternatives is governed by the criteria for an environmentally sound disposal plan. These criteria are presently addressing the potential for surface or ground water pollution by either runoff or leachate percolation and the requirements for land reclamation of retired disposal sites. Whereas these criteria are established by state regulatory agencies and in as much as the disposal requirements are not yet fully defined by any state, it will be presumed that all disposal alternatives herein discussed may be equally acceptable.

In each of the disposal plans, the basic concept of control is containment or isolation. The extent or thoroughness of containment required will depend on the hazard the disposal is adjudged to present to the environment. For example, a sludge from which chemical pollutants may leach

may require containment within a disposal site with an impervious liner made of either natural or man-made materials. The most obvious example would be disposal in a clay or plastic lined pond. An alternative to this approach would be to reduce leaching by one of several commercially available chemical fixation processes. An alternative to chemical fixation may be direct land filling of dewatered sludge by methods that restrict rain water percolation and leaching. The acceptability of the latter method may depend on the geological and hydrological conditions of the disposal site. Variations of these methods and the advantages and disadvantages of each are discussed in the following sections.

5.1.1 Chemical Fixation/Landfill

To date, three companies - Chemfix, Pittsburgh, Pennsylvania; Dravo Corporation, Pittsburgh, Pennsylvania; and IU Conversion Systems, Inc., Philadelphia, Pennsylvania - have developed a significant technology for chemical fixation of power plant sludge materials at disposal costs that may be considered generally acceptable. Whereas all three processes are designed to produce a material that chemically binds soluble components, their processes and products are distinctly different. Although not all of these processes are currently being operationally demonstrated with power plant sludges, each processor has sufficient experience in the fixation of sludge materials to give credence to his performance claims. The following discussions reflect these claims, which are now being assessed by independent analyses in the Aerospace Corporation sludge disposal study.

5.1.1.1 Chemfix Process. Chemfix has noted^{7,8} that their company has gained wide field experience in the fixation of solid, sludge, and liquid wastes produced by automotive, chemical, mining, paint, petrochemical, steel and metals industries. The Chemfix process reportedly can handle sludge fixation over a broad range of percent solids and produces a product with a soil-like appearance which does not prevent the percolation of rain water, but does have structurally stable properties and chemically binds the constituents to accomplish pollution control.

The Chemfix process involves the reaction of sodium silicate and one or a combination of setting agents--portland cement, lime, calcium sulfate, calcium chloride--with the waste material to form chemically and mechan-

ically stable solids. The reaction occurs at ambient temperatures and is not affected by temperature variations. The particular chemical choice, chemical ratios, and reagent quantities depend on the type of waste, required speed of reaction, and the end use of the fixed material. Either reactive or nonreactive wastes can be treated with the Chemfix process. The reaction process involves an initial gelatinous state and a subsequent hardening period. The gelation time, which is controlled, is based on the distance the treated material must be pumped to the disposal site.

The treated waste can be made either hard or soft with varying textures depending on its ultimate disposal requirements. The chemical system reacts with all polyvalent metal ions producing stable, insoluble inorganic compounds. The resulting material reportedly can be used directly as landfill which, when properly fertilized, will readily support grasses and plant life.

Chemfix data further indicate that the material produced by the Chemfix process does not leach chemical constituents by rain water at concentrations that exceed the natural background levels in ground waters. Its soil-like behavior should allow it to be compacted such that land reclamation costs can be minimized for any subsequent use of the land. Chemfix claims chemically fixed sludges by this process should be acceptable to regulatory agencies for disposal directly as landfill without further treatment.

5.1.1.2 Dravo Process. The Dravo Corporation process is now being used by Duquesne Light Company to stabilize the sludge produced by the lime scrubbers at Duquesne's Phillips Station near Pittsburgh, Pennsylvania. Dravo claims⁹ that they have gained extensive laboratory experience on sludges from various air pollution control devices that include but are not limited to power plant sulfur dioxide scrubbers. This process can chemically fix sludges at various solids content and it produces a material that is clay-like in consistency.

The Dravo process chemically fixes sludge by the inclusion of a proprietary admixture called Calcilox. The Calcilox is mixed with the sludge usually just before the sludge is pumped to a disposal basin. The sludge solids are allowed to settle and curing takes place under water over a 30-day period. Disposal can be made directly in the final disposal site, or an interim curing basin can be used from which the cured solids can be removed to a permanent disposal site.

Dravo information indicates that the product of the Dravo process is a firm material which is convenient for disposal. By adding 3 percent Calcilox (based on sludge solids) the conditioned sludge has properties that compare with silty clay with permeability of 2×10^{-5} cm/sec. Clay-like properties including high shear and compressive strength are being obtained with Calcilox additions of 5 to 10 percent.

The low permeability, low leachability and the structural stability of sludges conditioned by the Dravo process should allow them to be used directly as landfill materials. The company claims that sludges chemically fixed by this process have been accepted for disposal without containment in at least one site in Pennsylvania.

5.1.1.3 IUCS Process. The IUCS process^{10, 11, 12} has been field tested on sludge produced by wet scrubbing effluent gases on a lead smelter using hydrated lime as the absorbent. This sludge has many properties similar to power plant sludges and produces a hardened composition when chemically fixed. Laboratory tests have verified the similarity of sludge fixation behavior between this sludge and power plant sludges.

The IUCS process uses fly ash and a lime additive as ingredients in the chemical fixation of sulfite/sulfate sludges. The quantity of either of these additives depend on the moisture content of the sludge and the reactivity of the fly ash. In some cases, dewatering of the sludge is necessary for the desired reactions to take place at economical amounts of the additive. Three primary reactions take place in this process: 1) the reaction of lime with soluble sulfate that originates in either of the fly ash or sludge liquor so as to form calcium sulfate (sulfite will also react, but more slowly), 2) the reaction of lime, sulfate and iron or

aluminum oxides present in the fly ash glass forms complex sulfo-ferrites or sulfo-aluminates (this reaction results in the formation of the crystalline phase ettringite which contains substantial quantities of water within the crystalline structure), 3) the reaction of lime with the glassy silica of fly ash results in the well known pozzolanic reaction that proceeds slowly to form the calcium silicate phase, tobermorite.

Some IUCS laboratory results of this process are shown in Table 4 for the Shawnee limestone sludge. The data indicate that the material can develop strengths as high as 4.1 tons/sq ft in seven days. Falling head permeability data indicate improvement during the first month of curing from 10^{-4} to 10^{-6} cm/sec. The low permeability is a consequence partly of the expansive nature of ettringite which tightly seals the set mass and which prevents shrinkage crack formation which may otherwise allow some leakage. Besides low leach rate, analysis of the leachate indicates a reduction in concentration of trace elements of approximately one order of magnitude. Thus, as a consequence of chemically combining trace elements into new crystalline phases and the reduction in leaching rate, the availability of soluble salts or toxic contaminants to ground waters can be reduced many times, when compared to unfixed sludge (see Tables 2, 3, 4). An independent analysis of this fixed sludge sample (as well as others using Calcilox and the Chemfix process) is now being conducted in the Aerospace Corporation laboratories, and further verifications will be made in an EPA field disposal program.

In addition to producing material suitable for landfill, the IUCS process may also be applied to the manufacture of synthetic aggregate having properties suitable for road base material. This application, which requires increased quantities of the additives and additional processing equipment (when compared to a landfill application), is currently being field demonstrated.

5.1.2 Dewatering/Landfill

One method of sludge disposal may incorporate dewatering, drying and placement of the waste solids by one of several techniques. The most direct technique would use the method of compaction developed for sanitary

TABLE 4. LABORATORY RESULTS OF FIXED TVA LIMESTONE SLUDGE ANALYSIS¹³

LEACHATE CONCENTRATIONS, ppm ^a	AGE OF TEST, days						
	2	4	6	10	14	28	
TOTAL ALKALI	1068	542 ⁻	810	524	40	40	
TOT. DIS. SOLIDS	1370	730	1210	770	180	250	
SO ₃	11	19	14	20	10	3	
SO ₄	45.2	41.9	36.2	51.0	48.5	43.6	
Cl	64	12	74	21	16	14	
Ca	268	220	235	170	27.5	20	
Mg	0.005	0.005	0.005	0.01	--	--	
Al	2.2	3.2	11.4	0.95	0.05	0.1	
Fe	0.02	0.01	0.01	0.05	0.01	< 0.01	
Mn	0.005	0.01	0.007	0.007	0.005	< 0.005	
Cu	0.02	0.01	0.01	0.02	0.005	< 0.01	
Zn	0.005	0.01	0.005	0.005	0.005	< 0.005	
Cd	0.007	0.005	0.005	0.005	0.005	< 0.005	
Cr ⁺³	0.05	0.03	0.02	0.02	0.01	< 0.01	
As	0.01	0.01	--	--	--	--	
Pb	0.2	0.05	0.05	0.05	0.05	< 0.03	
Sn	0.1	0.1	0.05	0.05	0.05	< 0.1	
Hg	0.01	0.01	--	--	--	--	
pH	12.35	12.5	12.1	11.9	10.2	9.0	

	AGE OF TEST, days			
	0	7	14	21
PERMEABILITY, cm/sec	1 x 10 ⁻⁴	6.3 x 10 ⁻⁵	--	6.2 x 10 ⁻⁶
COMPRESSIVE STRENGTH, tons/sq ft	--	4.1	4.7	--

^aEquivalent to mg/l

landfill. For the problem of leachability, containment in clay cells may be necessary. In some cases, final covering with clay may be all that would be necessary to prevent contamination of ground waters. The specific method of solid waste disposal will depend strongly on site requirements.

In most cases, partial drying is required to produce a material that is readily compacted so that the disposal site may be ready for structural use immediately after filling. The extent of drying required would be dependent upon the physical nature of the sludge. Several alternatives exist and their cost effectiveness depends on sludge characteristics, climatic conditions, and system variables. For a sulfate sludge dewatered by filtration or centrifugation, if any additional drying is required to reduce the water content to a level suitable for compaction, solar drying may be sufficient. On the other hand, for a sludge with relatively high sulfite content, dewatering by either filtration or centrifugation is not adequate for compaction. In this case, one alternative might include the blending of dry collected fly ash to sludge dewatered by filtration or centrifugation to reach optimum moisture content for compaction. A second alternative might employ a thermal dryer that uses waste heat to dehydrate the sludge. Depending on the many system variables, it is possible to pass all the dewatered sludge solids through the dryer to reach an optimum compaction consistency, or to pass through the dryer, a portion which can then be mixed with the remaining solids to reach proper consistency for placement. To acquire the proper degree of dryness, the two alternatives cited above may also be applied to the sulfate sludge. A third alternative would be to oxidize the calcium sulfite to gypsum and thereby gain the dewatering advantages of sulfate sludges.

The environmental acceptance of direct landfilling with dewatered sludge will be dependent on its potential for environmental pollution by leaching. The potential for leaching possibly can be minimized for sulfate sludge by cake washing and treatment of the wash liquor. However, this has not yet been demonstrated. For most cases, the cost for containment (e.g., with clay), if necessary, would equal or exceed the cost required for direct ponding in clay or an equivalent containment material. However, in the case of dewatered sludge disposal, the benefits of reducing the

disposal volume and the future land use of a compacted sludge site may offset the dewatering costs.

5.1.3 Environmentally Controlled Ponding

A potential disposal alternative is the ponding of raw sludge; e.g., clarifier underflow or dewatered sludge cake. In this concept, the sludge is transported (pumped or hauled) from the scrubbing operations to a waste disposal pond suitably constructed to contain the raw sludge for long periods of time. The pond configuration commonly used is essentially an excavated pit or basin with the soil graded into embankments or dikes generally 3 to 15 m (10 to 50 ft) in height. The pond bottom is usually flat, but may be contoured to drain off seepage. Impervious liner materials are installed along the inner pond surface to inhibit any seepage of liquids from the sludge into the ground. Both flexible (e.g., polyethylene, polyvinyl chloride, butyl rubber, chlorinated polyethylene, duPont Hypalon films) and nonflexible liners such as bituminous concrete, clay, and pozzolan stabilized base are used.

Leakage detection concepts are usually incorporated into the pond design. These concepts vary from a simple underdrainage basin with visual monitoring to an electrical system designed to measure the electrical resistivity of the ground in the pond area. In addition, many installations have visual or remote sensing systems to monitor the nearby surface and ground waters to determine if there is any leakage or seepage from the pond. Overdrainage can return pond surface water to the scrubber if appropriate to the total system design.

Although the basic design features of ponds are generally uniform throughout the country, there are local ordinances and environmental regulations that govern. Even though the concept of ponding is a proven technique for large-scale, low cost evaporation operations, the predicted useful life of a potential sludge disposal pond liner still needs to be defined. It is directly dependent upon the aging characteristics of the liner material and its chemical interaction with the sludge and the supporting soil. Flexible liners are generally guaranteed for 20 to 25 years¹⁴, and the life expectancy

of nonflexible liners is estimated to be somewhat longer. However, long term service data applicable to desulfurization sludge containment do not yet exist for either type of liner.

In addition, systems to control seepage from a sludge disposal pond after it has completed its service life have not yet been demonstrated. Such a system might consist of provisions for underdrainage and overdrainage combined with air drying to dewater the site, followed by capping with a contoured clay cover to provide for runoff of rain water. The purpose of such a system would be to eliminate a hydraulic head, thereby preventing seepage from the pond.

6.0 DISPOSAL COST ESTIMATES

Disposal cost estimates were made for each of the two most prevalent disposal methods: ponding, and landfill using chemical fixation. Information was provided by various sources, and for each method the cost data are widely variant. As a result, the disposal costs are given in ranges of values consistent with appropriate qualifications based on the many variable parameters attendant to either disposal method. Although refinements are expected during succeeding disposal cost studies in this program, the disposal costs are expected to be always expressed as a range of values, except when applied to a specific disposal operation. This section summarizes this analysis to date and includes significant disposal costs and appropriate rationale. All costs are given in units commonly used in industry, e.g., dollars/short ton (dollars/0.907 metric ton) and dollars/sq yd (dollars/0.836 sq m).

6.1 CHEMICAL FIXATION DISPOSAL COSTS

The disposal costs by chemical fixation are dependent upon many factors such as: the fixation process and the contractor-operating efficiency; sludge characteristics including water, ash, and chemical content; disposal site physical characteristics; distance from power plant to disposal site; sludge transport method; disposal site land costs and ownership; residual value of disposal site; and environmental monitoring. Additionally, the costs depend upon whether the power plant uses a disposal

contractor or does the work itself. To determine these costs, data were obtained from personal communications, visits, published cost data,^{12, 15, 16} brochures, and testimony from the National Power Plant Hearings on Scrubbers, October - November 1973. Sources of these data were: power companies - Commonwealth Edison Company and Duquesne Light Company; chemical fixation companies - Chemfix, Dravo Corporation, and IU Conversion Systems, Inc.; and the EPA Control Systems Laboratory. Some estimated disposal costs (converted to a 50 percent solids basis) quoted by the power companies and fixation companies are listed in Table 5.

These values, equated to sludges that are 50 percent solids, range from \$2.00 to \$9.00/ton. All relevant factors pertaining to the values given are not available; however, based on the data given, it appears that except for extreme cases (e.g., development or demonstration), the general range of costs will be approximately \$2.50 - \$5.00/ton of sludge on a 50 percent solids basis. Some additional costs may be incurred for soil cover and/or seeding after the disposal site is filled.

Previously published fixation cost data include various limitations on the applicability of the values, and do not include recent escalations in costs. These data, therefore, have been augmented through personal communications with several of the companies involved for the purpose of deriving a general cost estimate for fixation/disposal rather than to attempt a rating of one versus another since each application would be condition and site dependent. Sources which provided estimates for complete disposal systems (using quite different approaches) were Dravo Corporation which provides a disposal service, and two power companies performing their own disposal tasks. Appropriate Dravo data are taken from costing of a proposed disposal operation for the Bruce Mansfield Station, a new 1650 Mw plant, in western Pennsylvania. The costing data are based on considerable land acquisition and development costs and the pumping of the sludge (32-37 percent solids) approximately 8 miles to the disposal site where it is to be chemically treated prior to disposal. In this case, although capital costs are high, the large quantities of sludge place the cost per

TABLE 5. SLUDGE FIXATION COSTING ESTIMATES

(all values are condition and site dependent)

SOURCE	DOLLARS/TON (50% solids)	REMARKS
COMMONWEALTH EDISON COMPANY, WILL COUNTY STATION		
	8.00 - 9.00	CURRENT ESTIMATE ^a
	6.00 - 6.50	TARGET ^a
DUQUESNE LIGHT COMPANY, PHILLIPS STATION	~ 7.50	TOTAL DISPOSAL
IU CONVERSION SYSTEMS, INC.		
CHEMFIX		
DRAVO CORPORATION	2.00 - 6.00	TOTAL DISPOSAL

^aCompany operation - on-site disposal; excludes capital costs

ton as analyzed in this study at less than \$5.00. For sites requiring less land development and shorter pumping distances, Dravo estimates costs of less than \$4.00 per ton. On the other hand, Commonwealth Edison and Duquesne Light Company are currently conducting their own systems on what can be termed development operations. These are, by comparison to the Dravo example, relatively small systems (156 and 100 Mw, respectively), and they use intermediate curing ponds and trucking of the cured sludge to disposal sites up to 1.6 Km (1 mile) from the fixation operation. Estimated current costs for these two are in the range of \$7.50 to \$9.00 per ton on a 50 percent solids basis. Some estimates of power plant operation such as the one at Commonwealth Edison indicate that refinements in the system may bring the cost down to approximately \$6.00 per ton on an operational basis. Additional contacts with Chemfix and IUCS to up-date published information indicate that for fixation disposal of 50 percent solids sludge including all capital charges and local transport of the treated sludge, a cost of \$4.00 to \$5.00 per ton is reasonable.

6.2 PONDING ECONOMICS

The cost for constructing a sludge pond will depend not only on the technical design details, but it will also depend upon its geographical location. The cost for shipping the liner materials to the site from the factory or processing center will vary with the distances traveled; and, the labor costs for construction and liner installation vary with the different areas of the country. The economic analysis undertaken in this study considered these variations and used average values such that the costs presented in this paper can be considered typical, but not necessarily specific.

Cost estimates are from suppliers of pond materials and builders of ponds^{14, 17}. All costs are as of August 1974. The economic analyses are based on a 1000 Mw power plant burning coal with 3 percent sulfur and 12 percent ash. The plant life is 30 years with the plant operating an average of 4560 hours per year with an average annual sludge (50 percent solids including ash) production of 930,000 metric tons (1,025,000 short tons).

The annualized costs used in this work included the costs for labor, maintenance and capital charges at 18 percent. The capital charges included depreciation, replacements, insurance, cost of capital, and taxes.

Figure 5 presents a composite cost summary for the various liner materials considered. The values shown are typical of the range of costs that might be expected for installed liners at the field site, but are to be used for planning purposes only. For this study, the mid-point of the price range was used. The widespread variations in costs resulted because of factors such as the cost of transportation from the liner manufacturer's facility to the field site within a radius of 1609 Km (1000 miles), the variation in the cost of labor throughout the country for liner installation, and price discounts for purchasing large quantities of liner materials. It is expected that the flexible liners could probably be purchased and installed at the lowest cost values; however, these lower limit numbers were avoided in the cost analyses as a precautionary measure against data which may not be reliable.

In early 1973, it was reported that the projected cost for a 46-cm (18-in) thick clay liner installed in several large ponds being constructed in the midwest was about \$6.00 per sq yd. This clay was transported by truck about 48 Km (30 miles) from the clay pits directly to the pond site. If suitable clay could have been available on the pond site, it could have been moved into place at a cost of \$1.00 per sq yd. At today's prices, the costs have risen to approximately \$9.00 and \$1.50 per sq yd, respectively. The increase in installed costs during this same time period for the flexible liners has been about 100 percent for 0.254 mm (10 mil) thick polyethylene liners, 50 percent for 0.508 mm (20 mil) thick polyvinyl chloride, and 25 percent for 0.762 mm (30 mil) thick Hypalon. The other flexible liners also have had appreciable rate increases. It is expected that the prices will continue to increase in the foreseeable future, since many of the liner materials are presently in such short supply that the prices vary on a monthly basis. Some contractors have indicated that the actual price will be established only at the time of delivery to the field; they will probably also consider writing escalators or surcharges

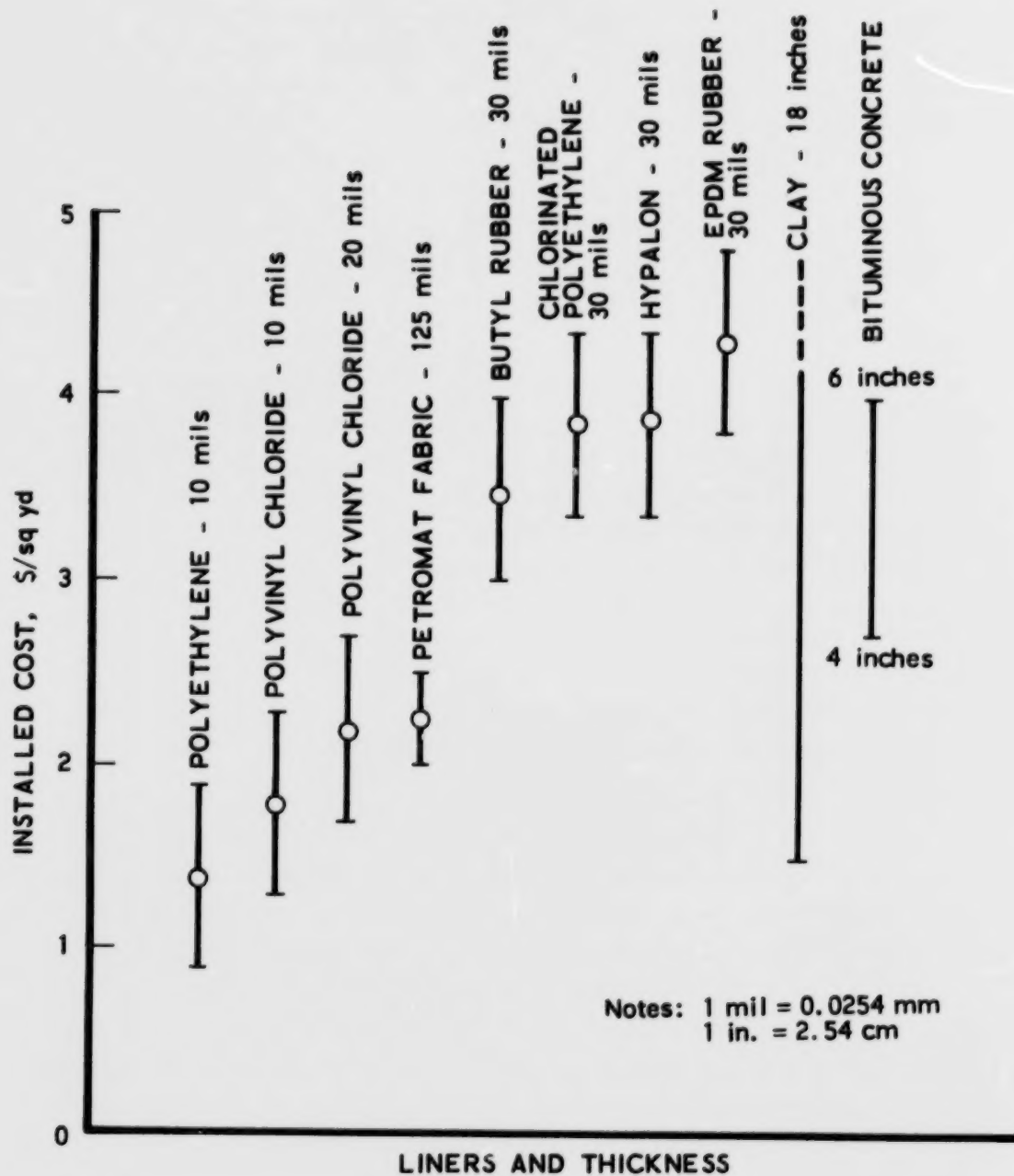


Figure 5. Installed liner costs

into the fixed price contracts to permit then to tag on price increases as the materials they purchase go up in costs.

Figure 6 illustrates the 30-year average disposal cost in dollars per ton of sludge (50 percent solids) as a function of the depth of raw sludge in the pond from 3 to 12 meters (10 to 40 ft). Two liner materials were considered: 0.508 mm (20 mil) thick polyvinyl chloride (PVC) at an installed cost of \$2.33/sq yd including a soil cover of 15.2 cm (6 in), and 0.762 mm (30 mil) thick duPont Hypalon at a cost of \$3.85/sq yd installed. These unit costs are also applicable to equivalent cost liners (see Figure 5). It can be seen that the disposal cost is (without including the cost of land) about \$2.50 per ton if PVC is used, and about \$3.40 per ton when a higher cost liner such as Hypalon is used.

Each pond configuration used in these analyses had a 0.915 meters (3 ft) freeboard. It was assumed that all of the land needed to contain the sludge during the full 30-year life of the power plant would be purchased initially. However, the sludge ponds would be constructed each 10 years as the previous ones became filled. Land costs of \$404 and \$2020/hectare (\$1000 and \$5000/acre) were used. As expected, the land cost decreased proportionately as the dike height increased because of a smaller area needed for the ponds to contain the 30-year production of sludge. Figure 6 illustrates the effect of the cost of land on the sludge disposal costs. It also indicates that for each pond concept there is an optimum sludge depth at about 9.15 meters (30 ft) which represents theoretical minimum disposal costs.

6.3 EFFECT OF DEWATERING

The disposal costs presented thus far were based on a desulfurization sludge with a mixture of ash and water at a total solids content of 50 percent. Over the 30-year operating life of the example power plant, sludge at this solids level will occupy a volume of about 2720 hectare-meters (22,000 acre-ft).

An investigation is underway to assess the effects (on pond size and disposal costs) of the dewatering of the sludge prior to disposal. Figure 7 graphically presents the results of the initial work comparing the volume occupied by:

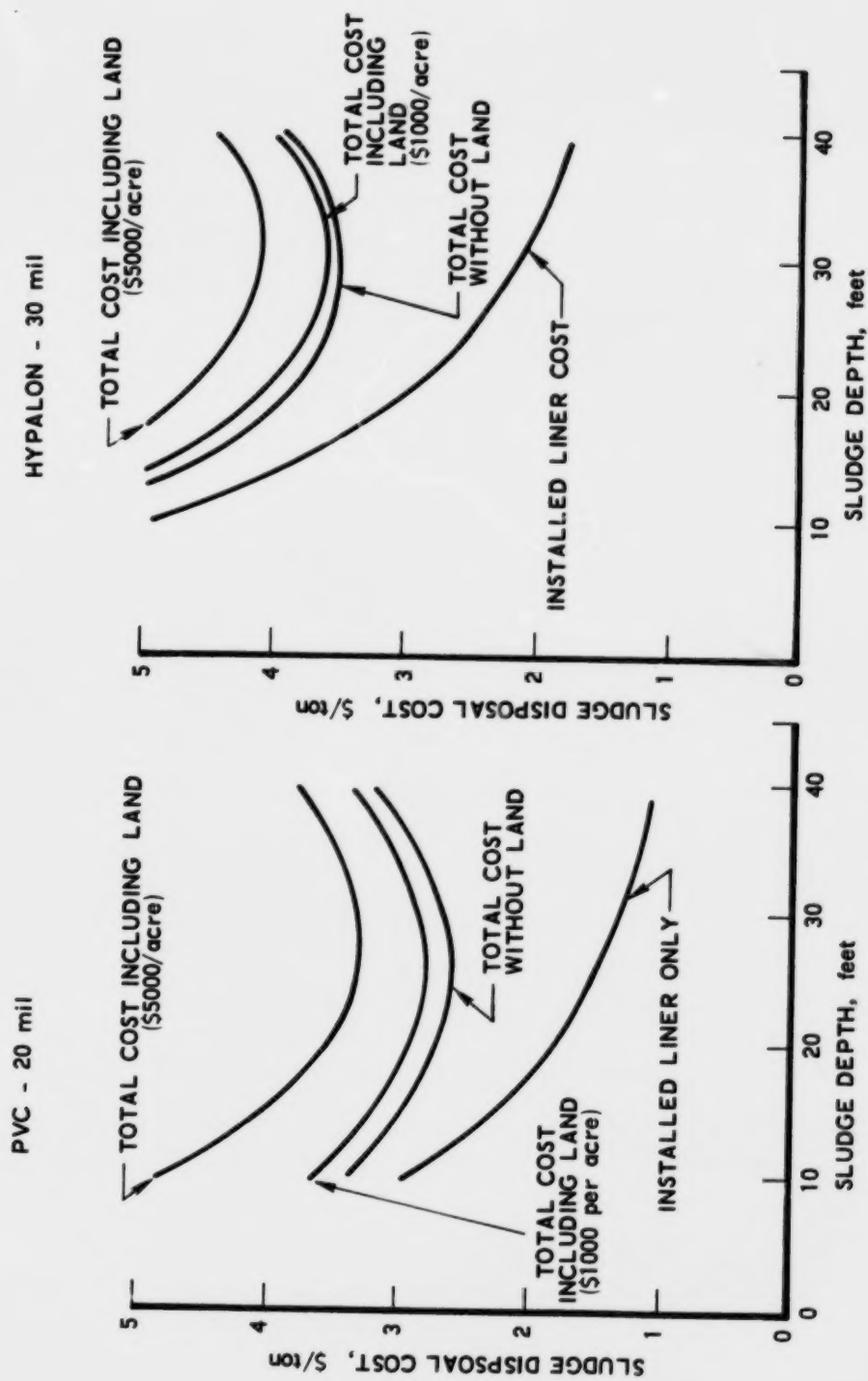


Figure 6. Disposal costs - ponding sludge, 50 percent solids (30 Year Average)

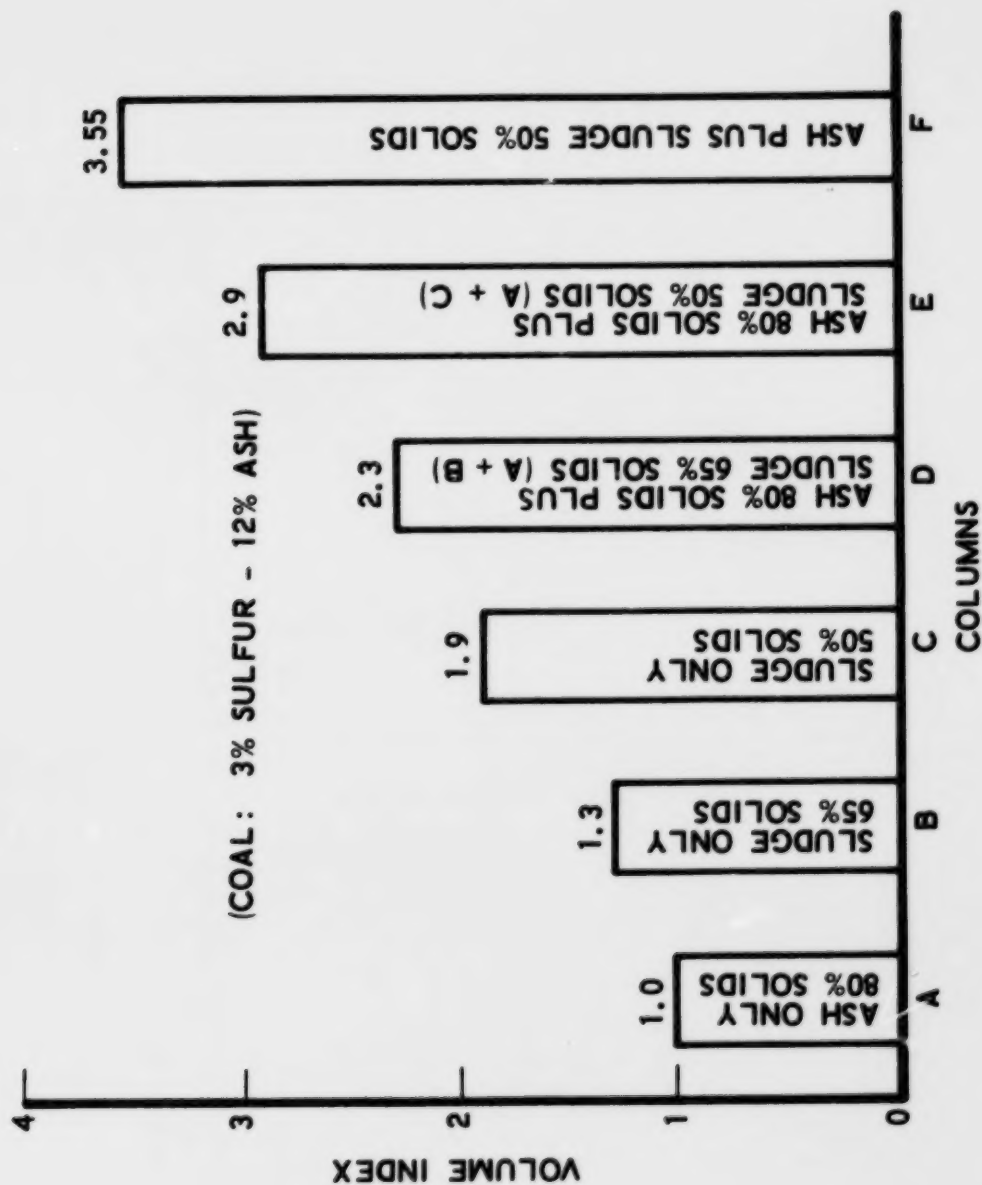


Figure 7. Effect of dewatering and separate ash collection on total sludge volume

only sludge at 50 percent solids, sludge at 65 percent solids, and then separately mixing these sludges with 80 percent solids ash. The reductions in volumes required for sludge disposal as a result of dewatering and/or ash removal prior to disposal are shown. For example, the disposal of sludge plus ash (50 percent solids) requires a pond volume 3.55 times that required for ash only; whereas, the disposal of sludge only, at 65 percent solids, requires a volume only 1.3 times that for ash alone. This concept in ponding offers the potential advantage of reducing the disposal costs by reducing the pond size and land required for the pond.

Economic trade-off evaluations are in progress to determine: 1) the various cost factors for conventional fly ash collection and sludge dewatering (at numerous solids content), 2) the disposal cost savings attributed to these factors, and 3) the net effects of 1) and 2). The dewatering techniques currently being considered include operations such as filtration, centrifugation, thermal drying, oxidation of the clarifier sludge, and settling in a thickener or settling pond.

6.4 DISPOSAL COST COMPARISONS

The disposal of scrubber sludges using a chemical fixation process is a rather new industry; therefore, little cost data exist based on actual practice by the power utilities. Similarly, little actual cost data exist for scrubber sludge disposal by lined ponding. However, the art and technology of ponding are well established, but there is little experience with power plant sludge disposal through the use of lined ponds or of subsequent pond reclamation. There are many ponding choices available that offer a wide range of costs with apparently small differences in net results. Therefore, at this time, comparisons are made by observing the ranges of values available and making rational comparisons in light of the variations in both the requirements and potential disposal methods.

It should be acknowledged that these two methods, which appear to be the best currently available or now in development, do not necessarily offer the same degree of environmentally sound disposal. For example, fixation,

which may be the more expensive process of the two, appears to offer a permanent solution and may be suitable for land reclamation, but further study is needed to evaluate the long-term environmental acceptability of this process. Flexible liner ponding, which may be the least expensive, may provide safeguards at a particular site for as long as approximately 25 years, but some added costs may occur for final disposal safeguards which may be necessary to provide permanent environmental protection and land reclamation. Although nonflexible liner systems may protect for periods greater than 25 years, eventually additional costs may also be required to accomplish permanent environmental protection and land reclamation.

Comparisons of fixation and lined ponding, therefore, cannot necessarily be made on an equal basis since the two methods do not necessarily offer the same results. If the two could offer the same environmental protection indefinitely and if land reclamation were not of primary interest at a particular disposal site, then a reasonably equal comparison could be made. Regardless of the many variations and differences that exist between the two processes, both are being considered throughout the power generating industry; therefore, cost comparisons are in order at this time.

Ponding costs are estimated to be in the range of \$2.50 to \$4.50 per ton depending on factors such as the moisture content of the disposal sludge, dike height, liner material used and the cost of land. Fixation cost estimates vary from \$2.00 to \$9.00 per ton depending on factors such as sludge chemistry including water content, chemical additives used, capital costs and transport method. Neglecting development factors and unusual circumstances, it appears that average ponding costs (as qualified previously) should be about half fixation disposal costs.

The following example considers the cost of disposal of all sludge and ash (50 percent solids) in terms of power produced; limestone scrubber; 85 percent SO_2 removal; 1.2 $\text{CaCO}_3/\text{SO}_2$ mole ratio; coal--3 percent sulfur and 12 percent ash--burned at the rate of 0.88 lb/kwhr. If the average disposal cost is in the range of \$2.50 to \$5.00 per ton, the cost for this example would be 0.56 to 1.12 mills/kwhr respectively. If a lime scrubber is used,

the cost of disposal would be about 10 percent less because the tonnage would be reduced. Additionally, reductions in disposal costs may be possible as noted in Section 6.3.

7.0 SHAWNEE SLUDGE DISPOSAL FIELD DEMONSTRATION

The Aerospace Corporation is participating in a scrubber sludge disposal field demonstration program being conducted by the EPA Control Systems Laboratory at the TVA Shawnee Power Station in Paducah, Kentucky. This program, which includes the monitoring of test ponds with each pond containing a raw or chemically conditioned lime or limestone sludge, will provide field condition data that will be integrated with the data being developed in the Aerospace Corporation disposal study. The Aerospace participation in the field demonstration consists of test planning, coordination, selected testing (soils, input sludges, leachate and ground water well samples, and conditioned sludge cores), and reporting. A complete description of the field demonstration is given in Reference 18.

8.0 REFERENCES

1. Rossoff, J. and R. C. Rossi, Aerospace Corporation, "Disposal of By-Products from Non-Regenerable Flue Gas Desulfurization Systems: Initial Report," EPA-650/2-74-037-a, May 1974.
2. Public Health Service Drinking Water Standards, Publication No. 956, U.S. Department of Health, Education and Welfare, Public Health Service, Washington 28, D.C., Revised 1962.
3. Water Pollution Regulations of Illinois, Pollution Control Board, State of Illinois, Springfield, Illinois, July 1973.
4. Proposed Criteria for Water Quality, U.S. Environmental Protection Agency, Washington, D.C., 20460, October 1973.
5. "Federal Water Pollution Control Act Amendments of 1972, 92nd Congress - Second Session," United States Code Congressional and Administrative News, No. 10, West Publishing Company, 1972.
6. Uniform Building Code, Table 29C, Vol. I, International Conference of Building Officials, 1970 Edition.
7. Personal Communication, Chemfix, Division of Environmental Sciences, Inc., Pittsburgh, Pennsylvania.
8. Conner, J.R., "Fixation/Solidification of Sludges from Lime/Limestone SO₂ Scrubbers"; Chemfix, Division of Environmental Sciences, Inc.^x, November 1973.
9. Personal Communications, J.G. Selmeczi, Dravo Corporation, Pittsburgh, Pennsylvania, February and May 1973.
10. Personal Communications, L.J. Minnick, IU Conversion Systems, Plymouth Meeting, Pennsylvania, March 1973.
11. Minnick, J.L., "Fixation and Disposal of Flue Gas Waste Products: Technical and Economic Assessment," Paper presented at Environmental Protection Agency Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973.
12. "Converting Stack Waste into Usable Products," POWER, L.J. Minnick, January 1974.
13. Personal Communication, R.G. Hilton, IU Conversion Systems, Inc., Plymouth Meeting, Pennsylvania, August 1974.
14. Personal Communication, O.H. Hensgen, Waterproofing Systems, Inc., Los Angeles, California, August 1974.

15. "Truck Loads of Landfill from Waste Sludge," Chemical Week, January 26, 1974.
16. Gifford, D.C., "Will County Unit 1 Limestone Wet Scrubber Waste Disposal". Presented at the Electric World Conference, Chicago, Illinois, October 30-31, 1973.
17. Personal Communication, W.A. Wheeler and H.C. Ellington, Jr., McKittrick Mud Company, Inc., Bakersfield, California, August 1974.
18. Jones, J.W., EPA Control Systems Laboratory (CSL) "Environmentally Acceptable Disposal of Flue Gas Desulfurization Sludges: The EPA Research and Development Program", for presentation at the EPA CSL Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.

AN OVERVIEW OF DOUBLE ALKALI PROCESSES
FOR FLUE GAS DESULFURIZATION

by

Norman Kaplan
Control Systems Laboratory
Office of Research and Development
Environmental Protection Agency
Research Triangle Park, North Carolina 27711

For Presentation At:

EPA Symposium on Flue Gas Desulfurization
Atlanta, Georgia
November 4-7, 1974

ABSTRACT

The chemistry and process design considerations applicable to sodium/calcium double alkali systems are presented. Technical terminology associated with these systems is defined.

The developmental efforts and full scale applications of the technology by Envirotech Corporation, FMC Corporation, General Motors Corporation, Zurn Industries, Arthur D. Little, Inc./Combustion Equipment Associates, Kawasaki Heavy Industries, Ltd./Kureha Chemical Industry Company and Showa Denko KK are discussed with reference to appropriate flow sheets. Planned applications of the technology by these companies are also discussed and tabulated.

ACKNOWLEDGEMENTS

The author wishes to express appreciation to Carolyn Fowler and Gloria Rigsbee for assistance in typing this paper, under the stress imposed by unusual time constraints.

Appreciation is also expressed toward the personnel listed below, without whose cooperation and informative input, a presentation of this type would not have been possible:

Conrad Cornell } Envirotech Corporation
William Ellison }

Jack Brady } FMC Corporation
Karl Legatski }

Thomas Dingo } General Motors Corporation
Thomas Mason }

John Tormey, Zurn Air Systems

Toshio Iihara } Kawasaki Heavy Industries
Makoto Yanai }

NOTES

1. Company Names and Products.

The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Consistency of Information.

The information presented in the section on Status of Technology was obtained from a variety of sources (sometimes by telephone conversation) including system vendors, users, EPA trip reports and other technical reports. As such, consistency of information on a particular system and consistency of information between the several systems discussed may be lacking. The information presented is basically that which was voluntarily submitted by developers and users with some interpretation by the author. The order of presentation of information or the amount of information presented for any one system should not be construed to favor or disfavor any particular system.

3. Units of Measure.

EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, NERC/RTP is providing conversion factors for the particular non-metric units used in the document. Generally, this paper uses British units of measure.

For conversion to the Metric system, use the following equivalents:

<u>British</u>	<u>Metric</u>
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft ²	0.0929 meters ²
1 ft ³	0.0283 meters ³
1 grain	0.0648 gram
1 in.	2.54 centimeters
1 in. ²	6.452 centimeters ²
1 in. ³	16.39 centimeters ³
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal.	3.7853 liters

TABLE OF CONTENTS

	<u>Page</u>
Introduction and Background	451
Definition and Discussion of Terms	453
Design Considerations	460
Status of Technology	475
Envirotech Corporation	476
FMC Corporation	484
General Motors	492
Zurn Industries	498
A. D. Little/Combustion Equipment Associates	500
Kawasaki/Kureha	501
Showa Denko KK/Ebara	506
Summary and Conclusions	509

INTRODUCTION AND BACKGROUND

Recently, there has been an increasing interest in the double alkali processes as alternatives to wet lime/limestone slurry scrubbing systems for utility and industrial boiler flue gas desulfurization (FGD) applications. Presently there are no full scale utility boiler double alkali applications in this country but there are two 150 Mw operating systems in Japan. In the U.S. there are two operating full scale industrial boiler applications equivalent in size to a 40 Mw and a 20 Mw control system respectively and an industrial kiln control system equivalent to a 10 Mw control system based on off-gas flow. The start up of a 20 Mw prototype utility boiler system is imminent. There has been a great deal of pilot plant testing of these systems in a variety of applications including coal and oil fired utility and industrial boilers, smelting plants, and in a sulfuric acid plant. There are now under construction in the U.S. (including the 20 Mw utility prototype) at least five full scale systems ranging in size from 3 to 150 Mw equivalent electric generating capacity. To put application of this technology in perspective, however, there are 11 utility boiler FGD applications now in service using lime/limestone scrubbing totaling over 2500 Mw of controlled capacity and at least 13 more systems under construction totaling over 7600 Mw of controlled capacity.¹

"Double alkali" or "dual alkali" processes as they have come to be known, like their precursors the lime or limestone wet scrubbing processes, are aqueous alkali scrubbing processes used for FGD. If a "black box" view of these processes is employed, they are, with minor exception, like the lime/limestone scrubbing systems: lime or limestone is consumed, and a calcium sulfite/sulfate and flyash wet solid waste product is produced. A more detailed examination inside the "black box", however, would show that the overall process has been split into a number of intermediate steps designed to improve upon the lime/limestone processes by increasing reliability of operation, utilization of lime and/or limestone, and sulfur oxide removal efficiency and, under certain circumstances, producing a solid waste with better handling characteristics. Whereas in a lime/limestone process the absorption of the SO_x from the flue gas and the production of the waste product occur to some extent simultaneously in a single reaction system, in the double alkali processes, these two steps are separated through the use of an intermediate soluble alkali; absorption and production of waste product can then occur in separate system components.

Separating the absorption and waste production functions accomplishes two very important objectives. First, it permits scrubbing the flue gas with a soluble alkali thus limiting the SO_x absorption reaction only to gas/liquid chemical equilibrium and to the rate of transfer of SO_x from the flue gas to the scrubbing solution. In lime/limestone processes, the rate of dissolution of lime or limestone is a third important limiting factor. Thus SO_x absorption efficiency in a double alkali system is potentially

higher than in a lime/limestone system with the same physical dimensions and liquid/gas flow rates. Additionally, soluble calcium is minimized and calcium slurry is kept out of the scrubbing apparatus thus preventing solids build-up (commonly known as scaling and plugging) in this critical area. Another benefit of this is better control of the lime/limestone reaction with acidic sulfur containing compounds in separate equipment specifically designed for this reaction thus potentially increasing lime/limestone utilization.

Although a number of other processes can technically be considered double alkali processes, this paper is limited to consideration of the sodium/calcium based double alkali processes. In these processes, a soluble sodium based alkali (NaOH , Na_2SO_3 , Na_2CO_3 , NaHCO_3) is used to absorb SO_x from the flue gas in the scrubber, and then a calcium based alkali (Ca(OH)_2 , CaO , CaCO_3) is reacted with the SO_x -rich scrubber effluent liquid to precipitate the insoluble $\text{CaSO}_3 \cdot \frac{1}{2} \text{H}_2\text{O}$ and $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ and regenerate sodium based soluble alkali for recycle to the scrubber system. Double alkali systems using ammonium/calcium base have been tested and while they might have advantages in the reaction with calcium compounds, their main disadvantage is the potential for pollution by a visible ammonium salt plume from the scrubbing apparatus caused by the highly volatile ammonium compounds in the scrubbing apparatus. Another variation that might be considered double alkali is Monsanto's "Calsox" process, which uses an aqueous organic base as the absorbent solution in combination with lime as the calcium supplying alkali to produce the throwaway product.

This paper will discuss some of the chemistry and the status of technology of sodium/calcium double alkali processes, concentrating for the most part on work done and data developed by Envirotech Corporation, FMC Corporation, Arthur D. Little, Inc., General Motors Corporation and Kawasaki Heavy Industries, Ltd. since these companies have made significant contributions to the development of the technology and are the front-runners in application of the technology.

DEFINITION AND DISCUSSION OF TERMS

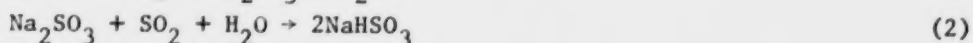
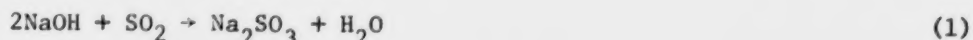
As with any specialized technology, a discussion of flue gas desulfurization in general, and double alkali technology more specifically, involves the use of special terminology which has evolved with the technology. While it is understandable to those dealing with the subject on a daily basis, it can be somewhat ambiguous to others. To clarify some of these ambiguities, and to define terms used here and by others describing double alkali technology development, a number of terms and concepts are defined and discussed in a general sense.

Absorption/Regeneration Chemistry

The main chemical reactions that take place in double alkali systems can be divided functionally into the absorption and regeneration reactions. A number of secondary reactions which have very important effects on the overall functioning of the system also take place. These include oxidation, softening and sulfate removal reactions which are discussed under the appropriate headings.

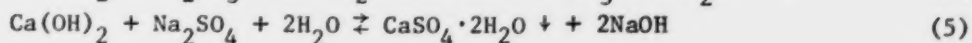
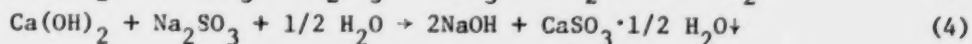
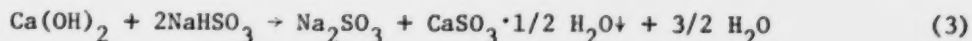
The regeneration reactions and in some cases the absorption reactions will be dependent upon which calcium supplying regenerant is used - lime or limestone. With lime the system can be operated over a wider pH range than with limestone. This wider pH range allows lime systems to operate over the complete range of active alkali hydroxide/sulfite/bisulfite whereas limestone systems can only operate in the sulfite/bisulfite range.

The main overall absorption reactions are described by the following equations:

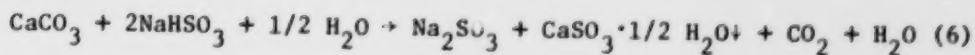


The main overall regeneration reactions are described by the following equations for lime and limestone respectively:

Lime

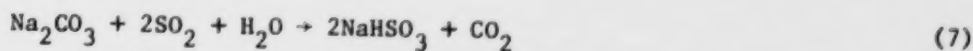


Limestone



Active Alkali

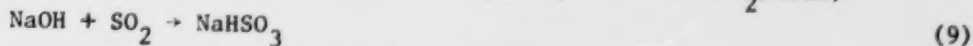
This term relates to concentration of NaOH, Na_2CO_3 , NaHCO_3 , Na_2SO_3 and NaHSO_3 in the scrubbing solutions. Sodium bisulfite is included in this definition although it is not technically an alkali (i.e., it cannot react with SO_2 in these systems); however, it can be converted to an alkali by reaction with lime or limestone. It should also be noted that the molar capacity of each of these species for absorption of SO_2 is different, and can vary from zero to 2 moles of SO_2 per mole of active alkali. This difference in molar capacity for absorption of SO_2 is illustrated by the following reaction equations:



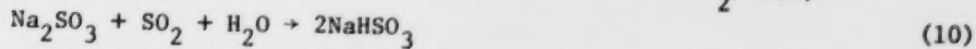
(Sodium carbonate molar capacity: 2 moles SO_2 /mole)



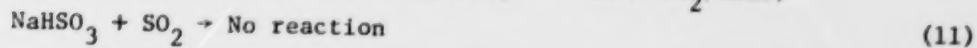
(Sodium bicarbonate molar capacity: 1 mole SO_2 /mole)



(Sodium hydroxide molar capacity: 1 mole SO_2 /mole)



(Sodium sulfite molar capacity: 1 mole SO_2 /mole)



(Sodium bisulfite molar capacity: zero mole SO_2 /mole)

Molar capacity is simply the number of moles of SO_2 needed to convert 1 mole of the absorbent alkali completely to sodium bisulfite. Since there is a difference in the molar capacity of different active alkali components to absorb SO_2 , active alkali is a descriptive rather than a quantitative term. If the concentration of each of the active alkali components (moles/liter) is known, however, the capacity of the scrubbing liquor to absorb SO_2 (moles of SO_2 /liter of solution) can be calculated as the sum of each of the active alkali component concentrations multiplied by their respective molar capacities as follows:

Scrubber liquor SO_2 capacity (moles/liter) =

$$2 [\text{Na}_2\text{CO}_3] + [\text{NaOH}] + [\text{NaHCO}_3] + [\text{Na}_2\text{SO}_3]$$

TOS

This is an abbreviation for "total oxidizable sulfur". It denotes the concentration of sulfur compounds in solution in which the sulfur is in the +4 oxidation state. Simply, this is the total concentration of sulfite plus bisulfite.

$$\text{TOS (moles/liter)} = [\text{SO}_3^{\equiv}] + [\text{HSO}_3^-]$$

Sulfate is not part of TOS, since the sulfur is in the +6 oxidation state in this specie. Sulfur dioxide dissolving in scrubbing solutions increases the TOS in solution.

Active Sodium

This is the concentration of sodium in solution which is associated with the active alkali.

$$[\text{Na}^+]_{\text{act}} = [\text{NaOH}] + 2 [\text{Na}_2\text{CO}_3] + [\text{NaHCO}_3] + [\text{NaHSO}_3] + 2 [\text{Na}_2\text{SO}_3]$$

If NaOH, NaHCO₃, Na₂CO₃, NaHSO₃ or Na₂SO₃ solids are added to double alkali solutions, the increase in sodium ion in solution is "active sodium". If Na₂SO₄ or NaCl, for example, is added, the increase in sodium ion is "inactive sodium". Active sodium is not increased by the dissolution of SO₂ in scrubber solutions.

Oxidation

Oxidation in a double alkali system refers to the conversion of TOS to sulfate by one of the following equations:



Simple oxidation of SO₂ to SO₃ in the flue gas is also considered oxidation in the double alkali system:



Oxidation in the system has the effect of changing active sodium to inactive sodium, or active alkali to inactive alkali.

Oxidation may occur in any part of the system: in the scrubber, the reaction vessels, or in the solids separation equipment. Rate of oxidation in the system is thought to be a function of rate of dissolution of oxygen, pH of the scrubbing solution, and impurities present in solution. Oxidation rate is thus affected by composition of

the scrubbing liquor (scrubbing liquors containing high concentrations of dissolved salts may absorb oxygen more slowly), oxygen content of the flue gas, impurities in the coal and lime or limestone, and the design of the equipment (the regeneration and solids separation sections of the system in particular can be designed to limit dissolution of oxygen).

Oxidation rate is expressed as a percentage and is calculated from an overall material balance on the system:

$$\text{Oxidation rate (\%)} = \frac{[\text{SO}_4^{2-} \text{ leaving the system (moles)}]}{[\text{Total sulfur collected (moles)}]} \times 100$$

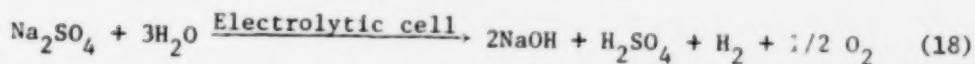
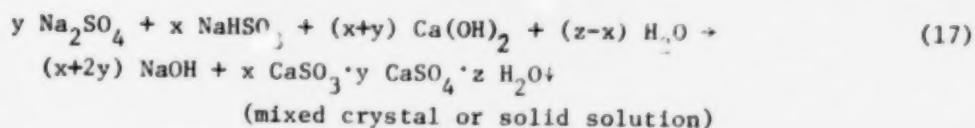
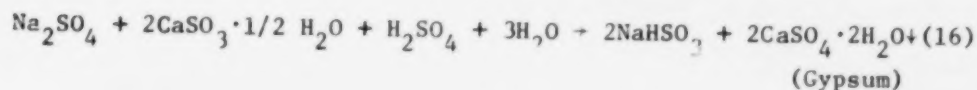
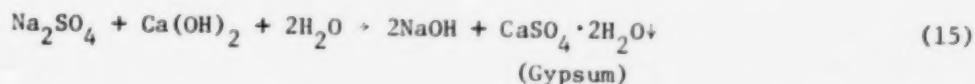
Sulfate leaving the system is total moles of sulfate in the solid waste plus any sulfate in the associated liquor.

Sulfate Regeneration

This term is a misnomer. What is really meant is sulfate removal from the system with regeneration of active alkali from inactive sodium sulfate. (See Sulfate Removal.)

Sulfate Removal

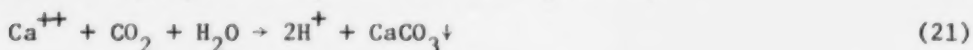
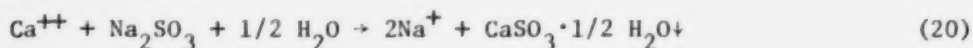
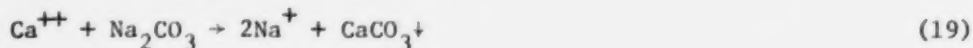
Sulfate is removed from the system with regeneration of active alkali from inactive sodium sulfate. Examples of these sulfate removal reactions are given below:



² Sulfate removal should be accomplished in an environmentally acceptable manner; a simple purge of soluble Na_2SO_4 from the system to land or water-way disposal is not acceptable in large quantities.

Softening

This term is used to describe various methods used to lower the dissolved calcium ion concentration in scrubber solutions. The purpose of softening the scrubbing liquor before recycling to the scrubber is to assure that it is subsaturated with respect to gypsum. This reduces the gypsum scaling potential in the scrubber. Following are examples of softening reactions:



In each of the above reactions, calcium ions are removed from solution as part of an insoluble material, outside the scrubber system. Reactions (19) and (21) are referred to as carbonate softening. Reaction (20) is considered sulfite softening. Generally, dilute systems employ carbonate softening while concentrated systems employ sulfite to prevent scaling in a system.

Dilute vs. Concentrated System

Dilute or concentrated refers to the active alkali concentration in a particular system. This differentiation is made because, in theory at least, based on their solubility products in water, both CaSO_4 and CaSO_3 should not precipitate from a solution of sulfite and sulfate simultaneously, unless the concentrations of sulfite and sulfate are present in a certain ratio. This can be shown by dividing one solubility product equation by the other:

$$\frac{[\text{Ca}^{++}] [\text{SO}_4^{=}] = K_{sp1}}{[\text{Ca}^{++}] [\text{SO}_3^{=}] = K_{sp2}}$$

from this, cancelling Ca^{++} ion concentration,

$$\frac{[\text{SO}_4^{=}]}{[\text{SO}_3^{=}]} = \text{constant}$$

The ratio of sulfate to sulfite for simultaneous precipitation of CaSO_4 and CaSO_3 is shown to be a constant.

The constant in the above equation is the ratio of solubility product constants of calcium sulfate and sulfite. Then, in theory, if

the ratio of sulfate to sulfite is higher than this constant, only calcium sulfate should precipitate; and if the ratio is lower than the constant, only calcium sulfite should precipitate.

This very simplified consideration of the chemistry given above is clouded in the "real world" by factors that contribute to non-ideal behavior in these systems. These factors include changes in ionic activities in solutions containing high electrolyte concentrations and evidence of coprecipitation of calcium sulfite and sulfate in the form of a "mixed crystal" or "solid solution" in a manner which is not completely understood at present.^{2,3}

With due consideration to the non-ideal behavior of these systems, however, under given conditions, a ratio of sulfate to sulfite in solution can be determined at which the previously cited examples hold. The ratio establishes a definition for "dilute" or "concentrated" double alkali systems. When the ratio is such that either gypsum or both gypsum and calcium sulfite will precipitate from the solution with the addition of slaked lime, then the system is "dilute".

Lime or Limestone Stoichiometry

Lime or limestone stoichiometry can be expressed as a percentage, based on an overall material balance around the system:

$$\text{Lime Stoichiometry} = \frac{\text{moles CaO added}}{\text{mole sulfur collected}} \times 100$$

$$\text{Limestone Stoichiometry} = \frac{\text{moles CaCO}_3 \text{ added}}{\text{mole sulfur collected}} \times 100$$

Lime or limestone stoichiometry is an indication of the efficiency of usage of lime or limestone in the system. Ignoring alkali components in the flyash collected, 100% stoichiometry is complete utilization; stoichiometries over 100% represent less efficient utilization of lime or limestone.

Feed Stoichiometry

This is calculated by a material balance around the scrubber. It is usually expressed as the ratio:

$$\text{Feed Stoich} = \frac{\text{Liquor SO}_2 \text{ Capacity (moles/liter)} \times \text{Flow (liters/min)}}{\text{SO}_2 \text{ (mole/min)}}$$

This ratio is evaluated for streams entering the scrubber.

Feed stoichiometry is a measure of the ability of the incoming liquor to react with or absorb all of the incoming SO_2 in the scrubber, assuming ideal contact of gas and liquor. Feed stoichiometry above 1.0 is required for high SO_2 removal. At feed stoichiometries at or below 1.0, assuming ideal contact between the gas and liquor, there will be significant equilibrium SO_2 partial pressure above the liquor, and thus SO_2 removal is theoretically limited to the value calculated on the basis of this SO_2 partial pressure in the exiting flue gas.

DESIGN CONSIDERATIONS

A commercial double alkali system must be designed to remove the desired quantity of sulfur oxides from a given flue gas stream, while operating in a reliable manner and discharging environmentally acceptable solid waste product. In fulfilling these design objectives, cost is also an important factor.

SO₂ Removal

The fact that small quantities of sulfur dioxide can be removed from large amounts of relatively inert gas by cyclic processes involving absorption into aqueous solutions of sodium sulfite/bisulfite has been known for some time. Johnstone et al.⁴ published a paper in 1938 giving data on the vapor pressure of SO₂ over solutions of sulfite/bisulfite and methods of calculating these equilibrium values under various conditions. The equilibrium partial pressure of SO₂ above sulfite/bisulfite solutions, the theoretical limit which a practical design can approach, is generally a function of solution temperature, pH, concentration of sulfite/bisulfite and total ionic strength. Since Johnstone's work a number of organizations have pursued this technology with laboratory, pilot plant and full scale applications for flue gas desulfurization, and many have demonstrated the ability for high removal efficiencies. (It should be noted that although Johnstone's work was aimed at cyclic processes with thermal regeneration, such as the Davy Powergas system, the vapor pressure data is also applicable to double alkali systems which use chemical regeneration.)

Once methods have been established to determine equilibrium SO₂ vapor pressure over scrubbing solutions, of the various concentrations to be encountered in an operating system, it becomes a matter of standard chemical engineering practice to design adequate gas absorption equipment to accomplish the desired SO₂ removal in a system. For comparison, it should be noted that the design of lime/limestone slurry absorption equipment is further complicated by the kinetics of dissolution of the lime or limestone, the particle size of the suspended material, and the crystal morphology of the lime or limestone.

Reliable Operation

System reliability can be adversely affected by two classes of problems, chemical and mechanical.

The mechanical problems include malfunction of instrumentation and mechanical and electrical equipment such as pumps, filters, centrifuges, and valves. These problems in a commercial FGD system can be minimized by careful selection of materials of construction and equipment and by providing spares for certain equipment such as pumps

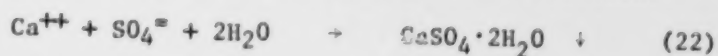
and motors which are expected to be in continuous operation and are prone to failure after a relatively short period of operation. Another important consideration in minimizing mechanical problems is the institution of a good preventive maintenance program.

The chemical (or physical/chemical) problems which may be associated with a double alkali system include scaling, soluble sulfate build-up, production of non-settling solid waste product, water balance and build-up of non-sulfur solubles which enter the system as impurities in the coal or lime. Each of these factors is associated with reliable system operation, or production of an environmentally acceptable solid waste.

a. Scaling - One of the primary reasons, and probably the most important, for development of double alkali processes was to circumvent the scaling problems associated with lime/limestone wet scrubbing systems. Therefore, a double alkali system should be designed to operate in a non-scaling manner.

Scaling is caused by precipitation of calcium compounds from scrubbing liquors, on the surfaces of various components of the system. When this occurs in the scrubber it is particularly troublesome since the flue gas path through the scrubber, if affected, could cause shutdown of the boiler-scrubber system and lower reliability.

Since scrubbing in double alkali systems employs a clear solution rather than a slurry, there is a tendency to ignore potential scaling problems. Testing experience with double alkali systems has indicated, however, that scaling can occur and indeed the problem should be a legitimate concern in the design of any system. Both gypsum and carbonate scale build-up has been recognized in these systems. Gypsum scaling is caused by the reaction of soluble calcium ion with sulfate ion formed in the system through oxidation of the absorbed sulfur dioxide or from absorbed sulfur trioxide according to the reaction:



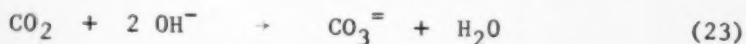
Gypsum scaling is controlled by softening the regenerated liquor prior to recycling to the scrubber. Softening ensures that the liquor recycled to the scrubber system is unsaturated with respect to gypsum; therefore, with proper softening even if some sulfate is formed in the scrubber, the liquor will not be saturated

with gypsum and cause scaling on the inside surfaces of the scrubber. In concentrated active alkali systems, a special softening step is not necessary since high sulfite concentration is maintained throughout the system. This sulfite maintains a low calcium ion concentration (sulfite softening), and thus maintains the scrubbing solution unsaturated with respect to gypsum.

Based on experience gained in lime/limestone scrubbing testing, a certain factor of safety in the prevention of gypsum scaling probably exists in double alkali systems. Gypsum has been found to supersaturate easily to about 130% saturation; thus, even if sulfate formation is higher than expected, gypsum may not precipitate in the scrubber until the liquor is over 130% of saturation with respect to gypsum.

Carbonate scaling usually occurs as a result of localized high pH scrubbing liquor in the scrubber where CO_2 can be absorbed from the flue gas to produce carbonate ion. This ion subsequently reacts with dissolved calcium to precipitate calcium carbonate scale according to the following series of reactions:

Carbon dioxide absorption by high pH liquor:



Calcium carbonate scaling:



Based on experience with the General Motors full scale double alkali system,⁵ carbonate scaling could occur with scrubber liquor pH's above 9. At lower pH, the carbonate/bicarbonate equilibrium system tends to limit the free carbonate ion and thus prevent precipitation of calcium carbonate:



Thus, carbonate scaling can be eliminated by control of pH in the scrubber.

b. Production of Non-Settling Solids - Under certain conditions, the waste product solids produced in the regeneration sections of various double alkali systems have a tendency not to settle from the scrubber liquors. This creates problems in the operation of settlers, clarifiers, reactor clarifiers, filters and centrifuges. Although

the phenomenon has been observed in the laboratory testing conducted by EPA on dilute systems and in the laboratory and pilot plant work conducted by Arthur D. Little, Inc. (ADL) on dilute and concentrated systems, it is not completely understood.³

Some of the factors thus far identified which appear to affect the solids settling properties are reactor configuration, concentration of soluble sulfate, concentration of soluble magnesium and iron in the liquor, concentration of suspended solids in the reaction zones, and use of lime vs. limestone for reaction. Based on laboratory work in dilute systems (about 0.1 M active sodium) using limestone it appears that solids settling characteristics degraded significantly at soluble sulfate levels above 0.5 M. Based on laboratory work with concentrated systems (about 0.45 M TOS, 5.4 pH, 0.6 M sulfate) using limestone, marked degradation of solids settling properties occurred at a magnesium level of 120 ppm and virtually no settling of solids occurred at the 200 ppm magnesium level. Equal degradation of solids settling properties also occurred in concentrated systems when the sulfate level was raised to the 1.0 M level while maintaining low magnesium level (about 20 ppm) and keeping other variables constant.³

Envirotech⁶ advocates the recycle of precipitated solids from the thickener underflow to the reaction zones in an effort to grow crystals which settle faster and are more easily filtered.

ADL cites reactor configuration as being important in the production of solids with good settling and filtration characteristics.³ Their basis for this is comparative tests of a simple continuous flow stirred tank reactor (CFSTR) with the ADL/Combustion Equipment Associates (ADL/CEA) designed reactor system under similar conditions. The ADL/CEA reactor system appeared to give better settling solids for a greater variety of conditions than a simple CFSTR.

Environmentally Acceptable Solid Waste

Ideally, the solid waste product produced by double alkali FGD systems should be environmentally acceptable. Some of the properties which can be ascribed to such a solid waste product are listed below:

- non toxic
- low soluble solids content, non-leachable
- low moisture content
- non-thixotropic
- high compressive or bearing strength

In order to generate waste product solids which have properties approaching those listed, the following design considerations are appropriate:

- sulfate removal
- water balance/waste product (cake) wash
- gypsum vs. calcium sulfite
- sludge fixation technology

a. Sulfate Removal - In double alkali systems, some of the sulfur removed from the flue gas takes the form of soluble sodium sulfate due to oxidation in the system, thus changing some of the active sodium to the inactive variety. When sodium in the system is converted to the inactive form (Na_2SO_4), it is relatively difficult to convert back to active sodium. To convert inactive sodium to active sodium, sulfate ion must be removed from the system in some manner, while leaving the sodium in solution. The alternative to this is to remove the sodium sulfate from the system at the rate it is being formed in the system. This alternative is not desirable since it is wasteful of sodium and generally is carried out by allowing the sodium sulfate to be purged from the system in the liquor which is occluded in the wet solid waste product.⁷ The solid waste product can then potentially contribute to water pollution due to leachability. Water run-off can lead to contamination of surface water, while leaching and percolation of the leachate into the soil can result in contamination of the ground water in the vicinity of the disposal site. Failure to allow for sulfate removal from double alkali systems will ultimately result in a) precipitation of sodium sulfate somewhere in the system if active sodium is made up to the system, or b) in the absence of make-up, eventual deterioration of the SO_2 removal capability due to the loss of active sodium from the system.

The equations (15), (16), (17) and (18) shown previously under the definition of "sulfate removal" describe several sulfate removal techniques which have been used in FGD system pilot tests.

The first equation depicts the sulfate removal technique used in dilute active alkali systems:



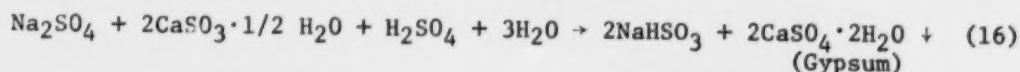
Concerning the full-scale dilute alkali system installed and operating at the Parma, Ohio, transmission plant of General Motors, and dilute systems in general, Phillips⁵ stated:

"The presence of Na_2SO_4 in the scrubber effluent is the prime factor influencing the design of the regeneration system. Na_2SO_4 is not easily regenerable into NaOH using lime. The reason being that the product, gypsum, is relatively soluble. . . . Na_2SO_4 cannot be causticized in the presence of appreciable amounts of $\text{SO}_3^{=}$ or OH^- because Ca^{++} levels are held below the CaSO_4 solubility product. To provide for sulfate causticization, the system must be operated at dilute OH^- concentrations below 0.14 molar. At the same time, $\text{SO}_4^{=}$ levels must be maintained in the system at sufficient levels to effect gypsum precipitation. . . . We selected 0.1 molar OH^- and 0.5 molar $\text{SO}_4^{=}$ as a design criteria."

In a previous paper, Phillips⁸ showed a plot of equilibrium caustic formation in $\text{Ca}(\text{OH})_2$ - Na_2SO_4 solutions at 120°F which is the basis for selection of the design criteria. The essence of this discussion is that if the active sodium concentration is sufficiently dilute, sulfate can be removed from the system by simple precipitation as gypsum by reaction of lime with sodium sulfate.

Since, as explained above, this reaction will not proceed to a great extent in concentrated active alkali systems, other techniques must be employed to effect sulfate removal in these systems.

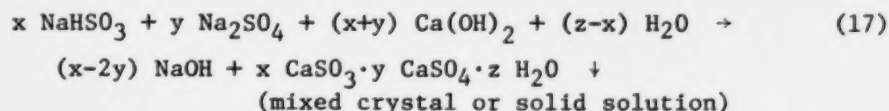
The second equation depicts a technique which is used in the full scale double alkali systems in Japan, and which has been pilot tested by ADL under contract with EPA:



This technique is used to precipitate gypsum by dissolving calcium sulfite in acidic solution thus increasing the Ca^{++} in solution enough to exceed the solubility product of gypsum. Ideally according to equation (16) 2 moles of gypsum should be precipitated for each mole of sulfuric acid added. In practice, however, this is not the case since any material which functions as a base can consume sulfuric acid and reduce the efficiency of this reaction for its intended purpose.³ Unreacted lime or limestone, sulfite ion and even sulfate ion can consume sulfuric acid thus lowering sulfate removal from the system.

Conceivably, this method of sulfate removal is economically unattractive in applications with very high oxidation rates, and where the gypsum produced must be discarded. The economic picture is considerably changed where this system is used merely as a slip-stream treatment to supplement other sulfate removal methods and/or where the solid product gypsum is saleable as is the case in Japan.

The third equation describes a phenomenon which has been referred to as mixed crystal or solid solution formation:



This phenomenon is described by R.H. Borgwardt of EPA² as it applies to lime/limestone wet scrubbing based on pilot plant investigations. A similar phenomenon has been observed by ADL in some of their early pilot testing of double alkali systems in conjunction with CEA, and later in the EPA/ADL dual alkali test program.

It appears that under certain conditons the solids precipitated in lime/limestone and double alkali systems contain sulfate, sulfite and calcium; however, the liquor from which these solids precipitated, appears to be subsaturated with respect to gypsum. This is based on the fact that pure gypsum crystal could be dissolved in the mother liquor from which the mixed crystal/solid solution was precipitated. In addition, the solid material was examined by X-ray diffraction and found to contain no gypsum; infrared analysis confirmed the presence of sulfate.

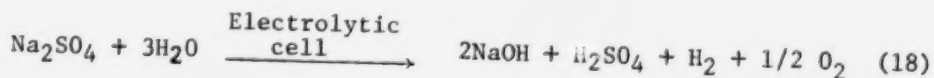
Borgwardt found that the molar ratio of sulfate to sulfite in these solids was primarily a direct function of sulfate ion activity in the mother liquor. In pilot test work with lime/limestone scrubbing, with little or no chlorides present and normal magnesium level (below 1000 ppm) in solution, the sulfate to sulfite molar ratio in the mixed crystal solids was found to reach a maximum level of 0.23. This is equivalent to a $[\text{SO}_4^{=}] / \text{total } [\text{SO}_x]$ ratio in the solids of 0.19.

In pilot test work with concentrated double alkali systems, ADL observed the simultaneous precipitation of sulfate and sulfite with calcium in lime and limestone treatment of concentrated double alkali scrubbing liquors. This phenomenon was surprising at first, in light of the reasoning which led to the development of dilute double alkali

systems; i.e., gypsum cannot be precipitated from solutions containing high active alkali concentrations. It was a simple technique for sulfate removal in concentrated systems. The $[SO_4]/$ total $[SO_x]$ ratio observed in pilot double alkali work was as high as 0.02 as a maximum.³ Coincidentally, this is the same value observed by Borgwardt in lime/limestone testing. This leads to the suspicion that the same phenomenon is occurring in both processes. The mother liquor from which these solids were precipitated was also found to be subsaturated in gypsum, and when the solids were examined, pure gypsum was not found.

Based on the observed data, it appears reasonable to design a concentrated active alkali system for a particular situation in which the system oxidation rate is below about 20%.³ In this case, sulfate can be removed at the desired rate, without the necessity for purging Na_2SO_4 or supplementing the system with other complex methods of sulfate removal.

The fourth equation shows sulfate removal as sulfuric acid in an electrolytic cell:



This method is the basis for operation of the Stone and Webster/Ionics process sulfate removal technique. In Japan, Kureha/Kawasaki has pilot tested the Yuasa/Ionics electrolytic process for sulfate removal in conjunction with their double alkali process. They feel that this process will be less expensive overall than the presently used sulfuric acid addition method. In addition, they feel that sodium losses from the system can be cut in half through the use of this method: from 0.018 moles Na loss/mole SO_2 absorbed to 0.009 moles Na loss/mole SO_2 absorbed.

Another approach to sulfate control is to limit oxidation. With sufficient limitation of oxidation, by process and equipment design, it may be possible to control sulfate by a small unavoidable purge of Na_2SO_4 with the solid waste product. To design for minimum oxidation, there should be minimum residence times in equipment where the scrubber liquor is in contact with oxygen-containing flue gas, and all reactors, mixers, and solids separation equipment should be designed to minimize absorption of oxygen from air. In addition, it has been reported⁹ that oxidation of scrubber liquors can be minimized by maintaining very high ionics strength. One possible

explanation for this is that high ionic strength liquors are poor oxygen absorbers and that oxidation in these systems is oxygen absorption rate limited.

b. Water Balance and Waste Product (Cake) Washing - In order to operate a closed system to avoid potential water pollution problems, system water balance is a primary concern. Water cannot be added to the system at a rate greater than the normal water losses from the system.

Generally there is a tendency to add fresh water to a scrubbing system to serve many purposes. These include:

- saturation of flue gas
- pump seal needs
- demister washing needs
- slurry make-up needs
- waste product washing

On the other hand water should only leave the system in the following ways:

- evaporation by the hot flue gas
- water occluded with solid waste product
- water of crystallization in solid waste product

Careful water management, part of which is the use of recycled rather than fresh water wherever possible, is necessary in order to operate a closed system.

As previously indicated, disposal of wet solid waste containing soluble salts is ecologically undesirable. In addition, allowing active alkali or sodium salts to escape from the system is an important operating cost factor. Sodium make-up to double alkali systems is usually accomplished by adding soda ash (recently quoted at \$49 per ton) at some point in the system. Thus, both ecological and economic considerations dictate that waste product washing is desirable.

A rotary drum filter, belt filter or centrifuge is usually the equipment in which the final solids separation is made. This equipment can be designed to permit solids washing with fresh water.

One concern in waste product washing is the extent to which the cake should be washed. At present, there are no federal regulations concerning the amount of total dissolved solids (TDS) which is permitted in a waste product which is to be disposed of as landfill. The future, however, may unfold stringent regulations in this area. One obvious consideration in waste product washing is system water balance. Unlimited waste product washing is not possible if a closed system operation with no liquid stream discharge is a goal. Another more subtle reason for limiting waste product washing is the potential problem of non-sulfur/calcium solubles build-up in the system. These non-sulfur/calcium solubles enter the system with the flyash, flue gas, the lime and/or limestone and the make-up water. Probably the soluble material in highest concentration would be sodium chloride which results from the absorption of HCl from the flue gas by the scrubber solution. A material balance around the system at steady state necessitates that solubles leave the system at the rate they enter. Thus, depending upon how well the waste product is washed, a certain level of non-sulfur solubles will be established in the system. Since the only mechanism for these solids to leave the system is as part of the wet solid waste, a certain purge rate is necessary. This purge also necessitates the loss of some sodium from the system. Practical limitations in filter design and water balance probably would limit a system to two or three "displacement washes" of the waste product (one displacement wash means washing with an amount of fresh water equivalent to the amount of water contained in the final wet waste product per unit of waste). Depending on the characteristics of the waste product and the design of the washing system, one displacement wash can reduce the solubles content of the waste product by as much as 80%.

c. Gypsum vs. Calcium Sulfite - Although gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$) is more soluble than calcium sulfite hemihydrate ($\text{CaSO}_3 \cdot 1/2 \text{H}_2\text{O}$), gypsum may be considered a more environmentally acceptable end product. The solubility of gypsum in water is about 0.25%; that of calcium sulfite is on the order of 0.0025%. It is interesting to note that while gypsum is a naturally occurring mineral, calcium sulfite is not found in nature. In the solid waste product (sludge) from lime/limestone and double alkali FGD systems, gypsum has better handling properties than calcium sulfite. Sludges containing a high ratio of gypsum to calcium sulfite are less thixotropic, better settling, more easily filtered, and can be more completely dewatered than sludges containing a high proportion of calcium sulfite. Another important characteristic which has been attributed to high gypsum (as opposed to calcium sulfite) sludges is their higher compressive or bearing strength. Typically, lime/limestone systems which generate

solids having a high proportion of calcium sulfite can be filtered to 40-50% solids, while some double alkali systems producing high gypsum cake can be filtered to over 65% solids.¹⁰

Some explanation for the behavior of these sludges is given by Selmeczi and Knight.¹¹ Although filter cakes appear dry, they still contain a considerable amount of water and thus, upon vibration or application of stress, they have the tendency of again becoming fluid. This thixotropic property and high moisture content are both explained by the morphology of calcium sulfite clusters. Due to the highly open, porous or sponge-like nature of these clusters, a considerable amount of water is retained in the clusters. The calcium sulfite crystals are rather fragile and break under pressure releasing some of the water, which results in the sludge becoming fluid.

It is possible in some double alkali systems to produce a high gypsum product. In Japan, where by-product gypsum is a saleable product, the calcium sulfite solids produced are oxidized completely to gypsum in a separate oxidation process tacked on to the tail end of the system. In applications where high excess combustion air is present in the boiler or where low sulfur coal is burned or a combination of these conditions is present, the oxidation rate in the system tends to be high (possibly about 90%) and the proportion of gypsum in the sludge tends to be high. In some dilute systems the proportion of gypsum in the sludge can be increased by augmental aeration of the scrubbing liquor.¹² Crystal seeding techniques used in conjunction with augmental aeration can produce relatively coarse grained gypsum crystals with good dewatering and structural properties in the final waste product.

d. Sludge Fixation Technology - Chemical or physical fixation of the sludge produced in a double alkali system is another potentially important means of producing an environmentally acceptable solid waste product. This technology is under investigation by I.U. Conversion Systems, Inc., Chicago Fly Ash, Dravo Corporation, and Chemfix Corporation. Most of their efforts are concentrated on sludge produced from the more prevalent lime/limestone systems; however, there is also some evaluation of double alkali sludges. The objective of sludge fixation technology is the production of a non-toxic, unleachable solid waste product which has reasonably high load bearing strength. If double alkali sludges are amenable to this type of treatment, the need to reduce soluble sulfates in the solid waste product becomes

less of a problem. Some sodium sulfate has been found to be physically or chemically tied up in the solid calcium sulfate/sulfite crystal lattice;³ however, the extent of this phenomenon is not generally considered to be adequate to remove all of the sodium sulfate produced via oxidation. Sludge fixation technology may be a mechanism by which additional sodium sulfate can be removed from the system without adverse environmental effects. There is some concern as to whether this is viable, however, since sludge fixation chemistry involves pozzolanic reactions between calcium compounds and flyash components in the sludge which may only involve multivalent ions rather than monovalent sodium. In other words, monovalent ions such as sodium may either a) not take part in the pozzolanic reactions, or b) inhibit or limit such reactions. Further investigation is called for in this area.

Dilute vs. Concentrated System

The selection of a dilute or concentrated double alkali system is an important design consideration in any application. In general it can be stated that the concentrated systems are more suited to applications in which oxidation is expected to be relatively low, and that, conversely, dilute systems are favored in applications where oxidation rates are high. High sulfur Eastern coal applications on utility boilers where excess air is controlled carefully and maintained at the lowest value consistent with complete combustion, the concentrated systems may be favored. On the other hand, in utility or industrial boiler applications where Western low sulfur coal is burned, and/or where control of oxidation is difficult due to high excess air, the dilute systems may be favored.

Oxidation rate is promoted when low sulfur coal is burned, since the ratio of oxygen to sulfur dioxide in the flue gas is higher than in high sulfur coal applications. Since oxidation is a strong function of the rate of absorption of oxygen, liquor which is dilute in TOS is subject to having a greater proportion of these species oxidized by a given amount of absorbed oxygen than one in which the TOS is more concentrated.

Under a given set of conditons without consideration given to waste disposal, a concentrated system can be installed at lower capital cost than a dilute system as previously discussed; however,

the desirability to produce a manageable solid waste (dilute systems can be designed to produce high gypsum sludges) could, in cases, override the economic issue.

Costs

Based on cursory design and cost analysis, with the assumption that a single scrubber device can be used to remove particulates and SO₂ to the extent required to meet new source performance standards, the capital and operating costs of double alkali systems appear to be significantly less than those for a lime or limestone system designed for the same requirements.¹³

Generally, the dilute active alkali systems tend to be higher in capital costs than the concentrated systems since both equipment size and process flows are required to be larger to accomplish the same degree of desulfurization. Methods used for sulfate removal and for reduction of scaling tendencies will affect both capital and operating costs.

Since there are presently no full scale utility applications in the U.S., a discussion of projected costs for such an application can only be based on costs for actual industrial boiler systems, vendor cost estimates and other engineering estimates based on hypothetical cases. Actual capital costs for industrial boiler applications tend to be high on a per KW basis compared to what would be expected in a large scale utility system because a) these systems are first-of-a-kind applications and consequently are designed for various operating contingencies, b) they are relatively small scale systems--economies of scale would be expected in larger scale units, and c) there are more than the usual equipment redundancies--e.g., four separate scrubbers to control only 32-40 Mw equivalent of flue gas. Vendor cost estimates tend to be less than complete and thus low, since they generally do not take into account all of the details that need to be considered for a specific application with its unique requirements. Engineering estimates are just that, estimates based on certain reasonable assumptions.

With the above qualifications, some double alkali system costs can be given.

a. Actual Industrial Boiler System⁵ - The General Motors industrial boiler system was installed for about \$3.5 million. Based on flue gas rate, it is presently equivalent to a 32 Mw system;

however, the regeneration system is designed to handle 40 Mw equivalent. The GM system is a dilute system.

Unit Capital Cost: \$88/Kw

Unusual features:

- 4 separate stainless steel scrubbers
- 2 separate filters
- overdesigned reactor clarifiers
- first-of-a-kind system

b. Vendor Cost Estimate - FMC Corporation¹⁴ has estimated the cost of a double alkali SO₂ removal process for a 150 Mw utility boiler system to be about \$6 million. Some of the assumptions and components of this estimated cost are shown in Table 1.

Unit Capital Cost Estimate: \$40/Kw

c. Engineering Cost Estimate - TVA has prepared a preliminary engineering cost estimate for a limestone double alkali process on a 500 Mw, new, coal fired power unit burning 3.5% sulfur coal designed for 90% SO₂ removal with on-site solids disposal.¹⁵

Unit Capital Cost Estimate: \$61/Kw

For comparison, on a similar basis, the TVA estimate for a limestone slurry process is \$50/Kw. Note that this comparison is in conflict with the cursory design referred to above.

Obviously the data base for double alkali system costs is extremely thin compared to that for lime/limestone systems. The costs cited above are given only to establish a basis for further consideration.

Table 1. ESTIMATED CAPITAL COST

FMC CORPORATION SO₂ CONTROL PROCESS

Retrofitted Parallel Scrubbers for 1-150 Mw Utility Boiler
and
Complete Regeneration Plant
Fully Instrumented, Installed & Operable

	Thousand \$		
	<u>Materials</u>	<u>Installation</u>	<u>Total</u>
<u>DIRECT COSTS</u>			
Scrubbing Section	497	430	927
Regeneration	381	706	1,087
Thickener and Offsites	465	310	775
Buildings, Foundations and Structural	<u>428</u>	<u>436</u>	<u>864</u>
SUB TOTAL A	1,771	1,882	<u>3,653</u>
<u>INDIRECT COSTS</u>			
Engineering (17,000 MH @\$16/hr.)			272
Field Indirects (140% of Direct Labor)			1,330
Fee (5% on \$5,255,000)			<u>263</u>
SUB TOTAL B			<u>1,865</u>
<u>OTHER COSTS</u>			
Proprietary Process Fee (Based on 150 Mw)			198
Contingencies (5% on \$5,518,000)			<u>276</u>
SUB TOTAL C			<u>474</u>
TOTAL A + B + C			5,992

NOTE: Includes chemical storage, reheaters, building, structural steel, flue gas ducting, waste solids handling and storage, insulation, steam tracing, startup, and operator training in addition to complete scrubbing system.

STATUS OF TECHNOLOGY

A number of variations of double alkali FGD systems have been tested and studied extensively during the past few years by equipment vendors, potential users, and EPA. Some of the results of testing and a description of some significant systems tested will be discussed here. In addition, some of the planned full scale installations will be discussed to illustrate the commitment made by industry to this technology.

In the United States, Arthur D. Little/Combustion Equipment Associates (the two companies have agreed to develop and market the technology jointly), Envirotech Corporation and FMC Corporation have all conducted extensive pilot plant testing with the aim of marketing sodium based double alkali systems for FGD applications on utility and industrial boilers. General Motors Corporation has undertaken to develop a double alkali process for use in their industrial boilers through pilot plant testing and, later, the installation of the first full-scale industrial boiler application of this technology in the country. General Motors does not intend to market their system; however, the results generated through their efforts will be in the public domain.

In Japan, the Showa Denko KK Company and Kureha Chemical Industry Company/Kawasaki Heavy Industries have developed sodium based double alkali systems through pilot plant testing and have gone on to relatively large full scale FGD applications which are currently operating. Due to different economics in Japan, the processes developed and installed by these companies produce saleable by-product gypsum rather than a disposable solid waste product. Another interesting difference between the development of technology in Japan and the U.S. is the use of limestone rather than lime as the source of calcium for the Japanese processes. In the U.S. the major emphasis in development and in commercial application has been with lime.

This discussion on the status of technology is presented by system developer or vendor, with emphasis on:

- System descriptions
- Test results/operating experience
- Plans for future testing or full-scale application.

Envirotech Corporation

The Envirotech testing is being conducted in a 2500 ACFM pilot plant at the Gadsby Station of Utah Power and Light Company in Salt Lake City, Utah. The pilot plant was constructed in late 1971. Testing began in January 1972. Flue gas is supplied to the pilot plant from No. 3 boiler, a 100 Mw coal fired unit.

a. System Description - Envirotech has tested and continues to test a number of configurations in the pilot plant. Particulate loading in the flue gas can be varied considerably since the flue gas take-off point from the power boiler can be upstream or downstream of a mechanical separator and an electrostatic precipitator or between the two. Two types of scrubber systems have been tested, a polysphere scrubber containing two trays of balls, and a venturi/cyclone separator scrubber. Two types of mist eliminators were tested, a spin vane in series with a cyclone, and a Euroform chevron type (this type of curved chevron demister allows the de-entrained liquid to flow from the system perpendicular to the gas path). Although the majority of testing was conducted in a dilute mode of operation which Envirotech calls the "sulfate mode" due to employment of augmental oxidation, they have recently also begun testing a concentrated mode of operation.

Envirotech has stressed equipment development in conjunction with process development. Use of reactor clarifiers is stressed to maintain high suspended solids content in the regeneration reaction zones. Quenching the incoming flue gas with recycle liquor (as opposed to fresh water) to avoid water balance problems has been stressed. Liquor introduction in a "dentist bowl" is used as a technique to avoid wet/dry interface fouling problems, while still providing the water needed to quench the hot flue gas.

Figure 1 is a slightly modified version of the basic sulfate mode flow scheme tested at the Gadsby station pilot plant by Envirotech. Although the majority of the testing was done with a polysphere scrubber, and the remainder with a venturi, this flow sheet incorporates both. Augmental oxidation is accomplished in a scrubber recycle tank by sparging with atmospheric air. (In the pilot plant over 90% oxidation was obtained with only a crudely designed sparging system; i.e., a rubber hose blowing air below the liquid surface.) This oxidation process converts most of the soluble sulfite to sulfate and thus the name, "sulfate mode". The purpose of this is to produce a waste product consisting mostly of gypsum and therefore having desirable settling, filtration and dewatering characteristics. This is workable, consistent with adequate sulfate removal, since the system operates in a dilute mode. Sodium sulfate is essentially converted to sodium hydroxide by lime precipitation of gypsum according to equation (15), with little interference of sulfite ion. The extent to which hydroxide can be regener-

ated in this way is limited to the range of approximately 0.1 molar OH^- concentration.

The scheme makes use of two reactor clarifiers, a thickener and continuous rotary drum vacuum (belt type) filter for solids separation. The reactor clarifiers are units designed for a combined job of reaction, clarification, and solid particle growth. Crystal growth and continued reaction are accomplished by internal circulation of suspended solids in the central zone of the equipment. The remainder of the equipment serves to split the feed slurry into a clarified liquor overflow and a thickened slurry underflow product. Reactor clarifier No. 1 serves these functions for the reaction tank effluent slurry. Reactor clarifier No. 2 is used for the softening step. Soda ash is added here serving the dual purposes of a) sodium make-up reagent and b) carbonate supplying reagent for softening. The soda ash reacts with soluble calcium, replacing some soluble calcium with soluble sodium, and precipitating calcium carbonate according to equation (19). This is carried out in the central reaction zone where the precipitated solids are caused to coagulate (grow) so that they may be more easily separated from the mother liquor. Clear liquor overflowing from reactor clarifier No. 2 is thus softened (subsaturated with respect to gypsum) and can be fed to the scrubber circuit as regenerated liquor with no gypsum scaling potential. It should be noted that reactor clarifier No. 2 was added to the pilot plant only after significant gypsum scaling was observed in the early testing in 1972. Coagulant, an organic chemical which may be added to the reactor clarifiers to the extent of a few parts per million, can serve to aid solids separation.

The thickener serves to make a gross separation of product solids from mother liquor which cuts down on the duty of the filter where the final separation and washing of the product solids occur.

The equipment arrangement used by Envirotech for testing a concentrated mode of operation is somewhat simpler compared to that used in the sulfate mode. The absorption section consists of a venturi/cyclone separator with recirculation. The regeneration system essentially consists of two series stirred tank reactors; the solids separation section consists of a single thickener and continuous rotary filter.

b. Test Results - General conditions and performance data for the Envirotech pilot plant operated in the sulfate mode are tabulated below:

Flue Gas

Source: pulverized coal boiler flue gas
Fuel: coal, 0.4% Sulfur
Oxygen content: 5%
Flow Rate: 1500-3000 ACFM @ 220°F
SO₂, inlet concentration: 250-1500 ppm
Particulate, inlet concentration: 0.2-0.3 gr/SCF

Absorber System

<u>Absorber Type:</u>	<u>Polysphere</u>	<u>Venturi</u>
SO ₂ Removal:	90%	80-85%
SO ₂ Exit concentration:	15-40 ppm	
Particulate exit concentration, gr/SCF:	0.01-0.02	
L/G ratio, gal./1000 ACFM:	20	24
Scrubber pressure drop, in. H ₂ O:	4-10	6-8
Recycle tank/oxidizer residence time, min.:	10-15	
Scrubber liquor --		
Inlet, pH:	7.5-7.7	
Exit (typical), pH:	5.8-7.1 (6.5)	
Oxidation rate, %:	~90	

Regeneration System

Lime stoichiometry: 110-115% of SO₂ removed
Reactor residence time: 30 minutes, total for both reactors
Regeneration liquor composition: 6.5 pH inlet
12.2-12.5 pH exit
0.1 Molar hydroxide, exit
40,000-45,000 ppm SO₄⁼
1000 ppm Cl⁻
~500 ppm Ca⁺⁺ (approximately 100 ppm below saturation of CaSO₄·2H₂O)

Solids Separation

Residence times: Reactor clarifiers: 1-2 hrs

Thickener: 3-4 hrs

Waste product composition: 30-40% water

1-2% flyash (dry basis)

2% solubles (dry basis)

10-15% unreacted CaCO_3 or

Ca(OH)_2 (dry basis)

remainder: CaSO_x solids

(mostly $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$)

Waste product wash: 1 displacement wash,
80% reduction in solubles

Discussion - The conditions under which the pilot plant operated, for the most part, simulated typical conditions for a utility boiler fired with Western low sulfur coal. The actual range of SO_2 concentration in the boiler flue gas was 250-400 ppm since coal containing approximately 0.4% sulfur was fired in the boiler. A much larger range of inlet SO_2 concentration is, however, indicated above, due to the fact SO_2 was added to the flue gas during some of the later testing to simulate conditions typical for a utility operating with a higher coal sulfur content. The majority of the testing with the higher SO_2 inlet levels was conducted with the venturi scrubber. During this testing the flue gas flow rate was at the lower end of the indicated operating range (~ 1500 ACFM).

The majority of the testing with the polysphere scrubber was conducted at a total pressure drop of 4 inches of water using two trays of balls. Figure 2 shows the performance of a single polysphere tray; SO_2 removal is plotted against L/G ratio in gal./1000 ACF at three different pressure drops (4, 7 and 10 in. H_2O) at pH in the range of 5.8-7.1. Figure 3 shows a plot of SO_2 removal vs. scrubber effluent pH at an L/G ratio of 19 gal./1000 ACF at a total pressure drop of 4 in. H_2O across two polysphere trays. It can be seen that, generally, two trays operating at a pressure drop of 4 in. H_2O are equivalent to one tray operating at a pressure drop of 7 in. H_2O .⁶

SO_2 and particulate levels achieved in the treated flue gas were well within the requirements of federal new source performance standards based on the data reported.

Lime stoichiometry reported for most of the testing was in the range of 110-115% based on SO_2 removed. Envirotech has forecasted, however, that utilization of lime could be improved so that in a commercial application one might expect to achieve lime stoichiometry in the range of 100-110% based on SO_2 removed.⁶

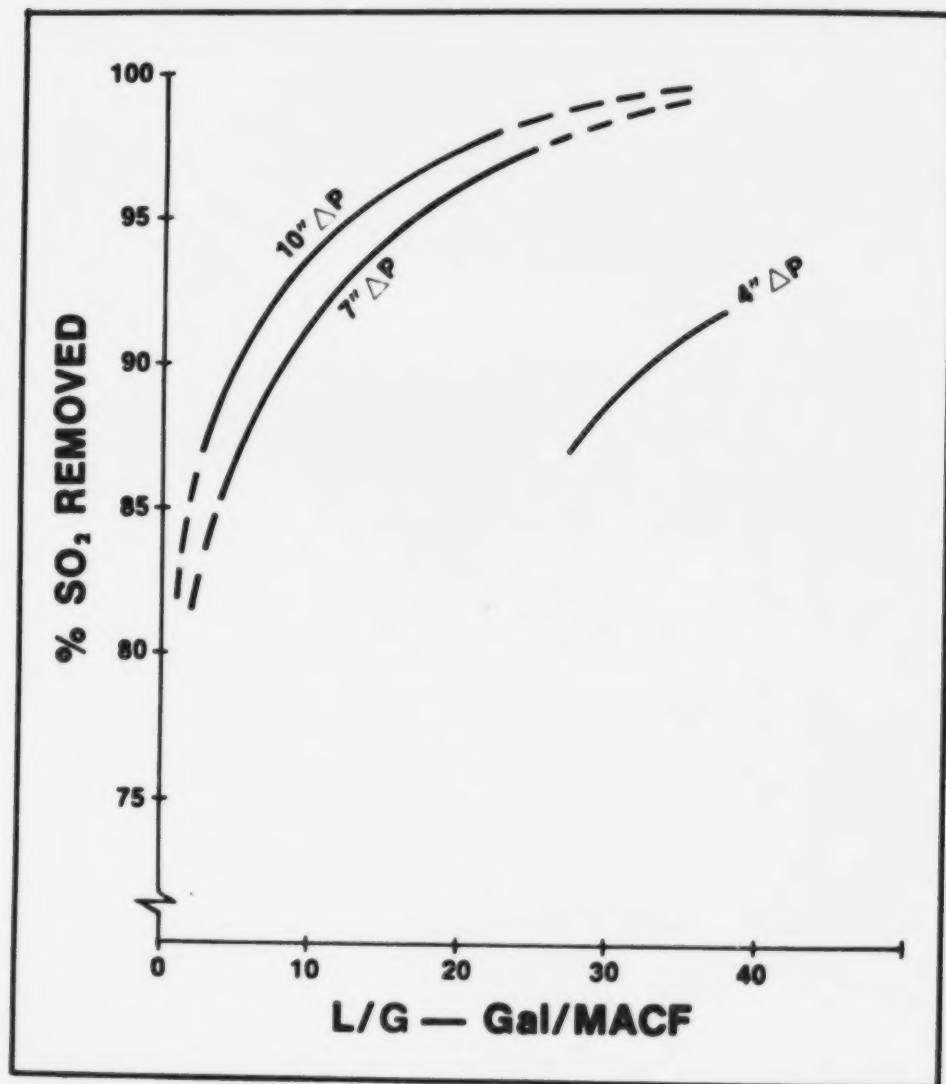


FIGURE 2. SO₂ removal vs. L/G. Coal fuel—electrostatic precipitator discharge. pH; 5.8 – 7.1, single-stage.

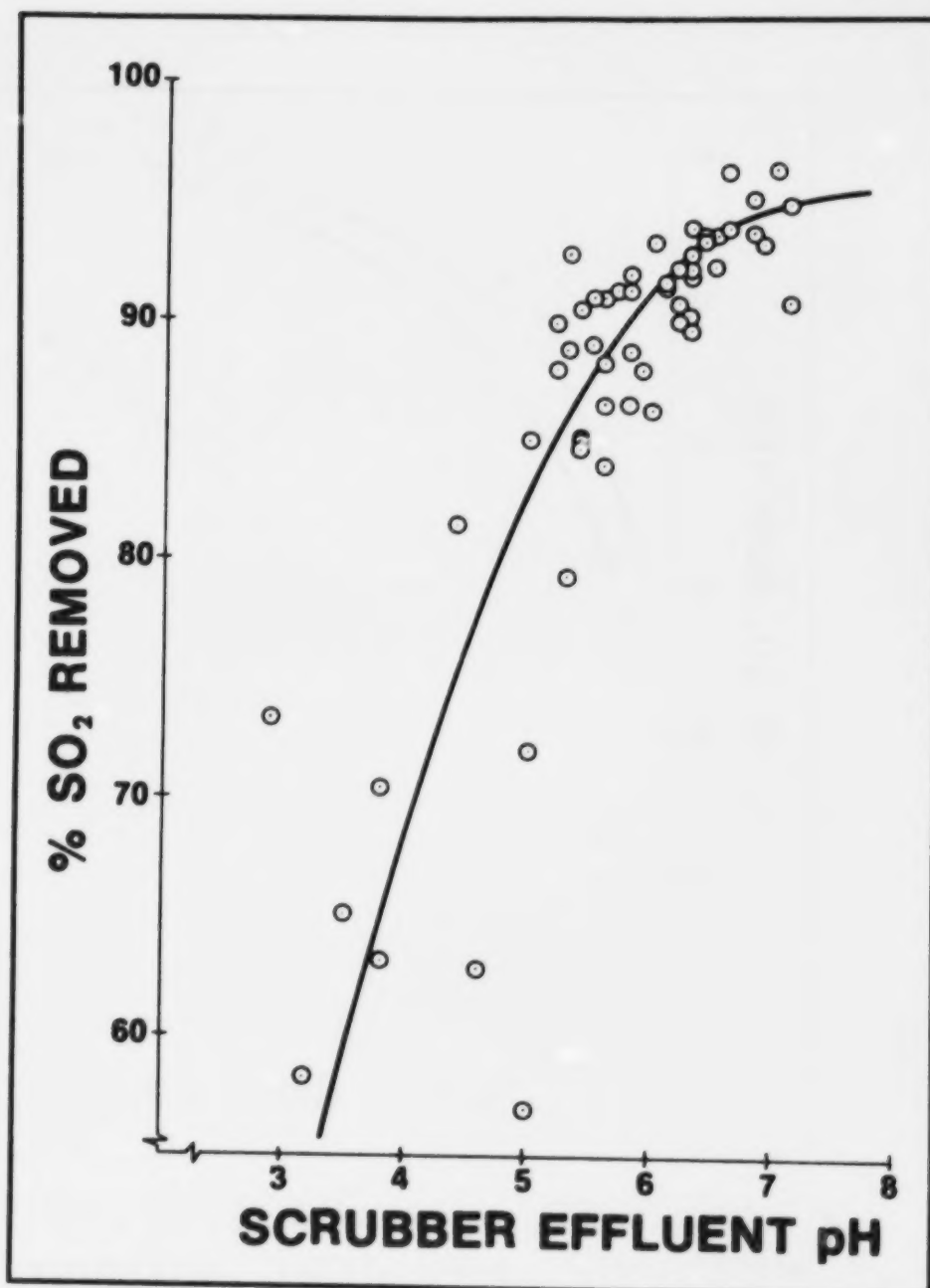


FIGURE 3. SO₂ removal vs. pH level. Coal fuel—electrostatic precipitator discharge. L/G, 19 gal/1,000 cu.ft., $\Delta P = 4$ in. H₂O. Two-stage operation.

The scrubber/regeneration liquors had a chloride content in the range of 1000 ppm at steady state. Soluble chloride is expected to concentrate in most wet scrubbing systems due to the conversion of coal chloride to HCl which appears in the flue gas and is absorbed by scrubber liquors. There were no apparent detrimental effects noted for chloride levels in the range experienced in this double alkali application.

The solid waste product, primarily gypsum, appeared to settle and dewater readily. The filter cake end product contained 60-70% solids, generally better than that produced in typical lime/limestone applications. The filter cake, however, contained about 2% solubles, somewhat higher than that expected in a typical lime/limestone system, but probably as low a value as has been obtained in steady state closed loop operation of a double alkali system. Experiments indicated that a single displacement wash could achieve as much as 80% reduction in soluble solids. The use of additional wash could conceivably lower the cake solubles content; however, in actual practice, the design of the filter, buildup of non-calcium/sulfur solubles in the system, applicable regulations concerning solids disposal and economic considerations would determine the need for additional washing.

Reliability - The pilot plant testing began in January 1972 and continued for 3 months during which period some gypsum scaling problems were encountered. The pilot unit was shut down for a period of about 5 months between the late spring and fall of 1972. During the 5 month shutdown a reactor clarifier was added to the flow scheme to accomplish the carbonate softening step which reportedly eliminated subsequent gypsum scaling problems.

The basic sulfate mode flow scheme was then operated for almost 2 years with no shutdown-causing, operating problems between fall of 1972 and August 1974. The pilot plant was operated mostly by a single technician per 8-hour shift for 24 hours per day, 5 days per week. The unit was shut down on weekends for convenience; however, there was no drainage of scrubbing or regeneration solutions or cleaning of equipment during the weekend shutdown periods. In November/December 1973 there was a brief shutdown period in which the venturi scrubber was tied into the system. After August 1974, testing of the sulfate mode was discontinued so that concentrated active alkali systems which Envirotech has termed "sulfite mode" could be tested and evaluated for comparison with their dilute sulfate mode of operation.

FMC Corporation

FMC began development work on a sulfur dioxide control system in the early 1960's at the Company's Modesto, California, chemical plant. In this plant, barium and strontium sulfate ores are processed in high temperature reduction kilns to produce barium and strontium sulfides with some evolution of SO_2 . After some initial experimentation with lime and limestone scrubbing processes in which severe scale formation problems were encountered, FMC began work on the double alkali process. They developed a concentrated active alkali scheme which was put into service controlling SO_2 emissions from two reduction kilns at the Modesto plant in December 1971. This was probably the first U.S. full-scale application of double alkali technology.

a. System Description - A typical schematic of the "FMC/Link Belt Alkaline Absorption Process" is shown in Figure 4.

General

Absorption - The scrubber system depicted here is what FMC refers to as their "dual throat" venturi and cyclonic separator. Absorption of SO_2 occurs by conversion of sulfite to bisulfite in the venturi (equation (2)). Due to the fact that FMC employs a very concentrated active alkali scrubbing liquor which is highly buffered by the sulfite/bisulfite equilibrium system, there is only a very small change in pH between the inlet and outlet of the venturi, on the order of 0.1 pH unit (e.g., typical operation of the FMC system might involve venturi inlet liquor pH of 6.5 and effluent pH of 6.4). Gypsum scaling does not occur in these systems since the calcium ion concentration is held well below that required for gypsum precipitation by the high sulfite concentration in solution. This can be referred to as sulfite softening (equation (20)). Carbonate scaling is not a problem in operation at these relatively low pH's due to the carbonate/bicarbonate equilibrium system which tends to depress the carbonate concentration and favor soluble bicarbonate.

Make-up sodium is fed as soda ash directly to the scrubber recirculation tank. No carbonate softening is intended in this step.

Regeneration - A relatively small bleed of scrubber liquor is taken from the scrubber recycle loop to the regeneration system. The regeneration system consists of a continuous flow stirred tank reactor with a 5 minute residence time and a thickener. Regeneration in this concentrated scheme involves only conversion of some of the bisulfite to sulfite by reaction with lime (equation (3)). There is no free hydroxide ion formation. Clarified liquor from the thickener rich in sulfite is fed back to the scrubber recirculation tank to raise the capacity of the scrubber liquor for absorption of SO_2 .

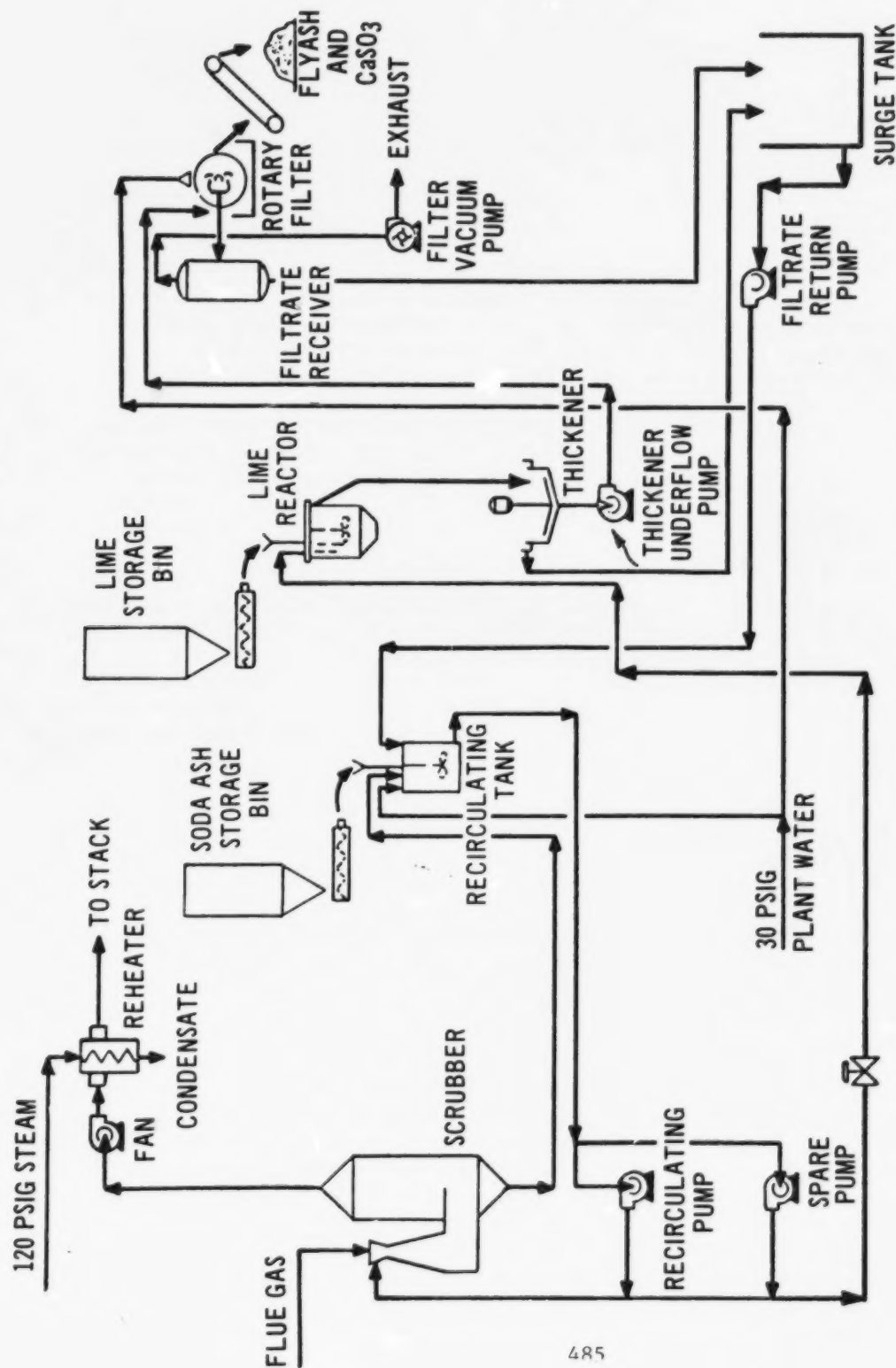


FIGURE 4. FMC/LINK-BELT ALKALINE ABSORPTION PROCESS FOR SULFUR DIOXIDE CONTROL
SIMPLIFIED FLOW SHEET

Solids Separation - Solids separation begins in the thickener which produces a concentrated slurry underflow for feed to the continuous rotary vacuum filter which makes the final separation of solids. The FMC system is designed for simultaneous removal of flyash and SO₂; therefore, the waste solids will contain a considerable amount of ash if the system is employed in a coal fired boiler application in which there is no separate provision for removal of flyash upstream of the control system.

Modesto System Description - The double alkali system in operation at the FMC Modesto chemical plant processes a combined 30,000 ACFM gas stream (550°F) containing 5000-8000 ppm SO₂ from two barium/strontium reduction kilns. The scrubber system consists of a tower (packed with 9 feet of Intalox saddles) and a recirculation tank. The regeneration system consists only of a single stirred tank reactor fed with lime and a bleedstream for the scrubber loop. Solids separation is accomplished in a rotary vacuum filter producing a solid waste containing 40-50% solids. There are no provisions for waste product washing. The overall system at Modesto is equivalent to 7-10 Mw based on exit flue gas flow, and possibly up to 35 Mw based on SO₂ removal capabilities.

b. Operating Experience and Testing - FMC has operating experience in the system at the Modesto plant since start-up in December 1971. In addition, FMC constructed a 3500 ACFM pilot plant in a mobile 40 foot trailer which has been tested at various locations on at least seven different flue gas and other waste gas streams for control of SO₂ and particulates. The basic flow scheme in the mobile pilot plant is that previously described and depicted in Figure 4.

Modesto System - This system has an enviable reliability and performance record since start-up in December 1971. As of early September 1974, the system had been on line for 22,800 hours at 100% availability for control of kiln off-gases. The kilns had scheduled shutdowns approximately every 3 or 4 months for maintenance and/or change of chemical process service; however, between December 1971 and September 1974, they were in service approximately 95% of the time. Operation of the system is now considered routine rather than experimental. With inlet SO₂ levels of 5000-8000 ppm, the system has achieved as high as 99.7% SO₂ removal; it routinely runs with less than 100 ppm in the exit off-gas stream.

The waste product solids from the system are discharged from a rotary vacuum filter at about 40-50% solids but there is no waste product washing. The discharged solids are reslurried to about 20-25% solids with other plant waste water streams to be sluiced out to an on-site disposal pond. The sludge dewatered by solar evaporation and percolation into the soil surrounding the pond. About once a year some

of the drier sludge is dredged from the pond for disposal elsewhere to create more on-site "running room".

The only serious drawback to the system is the large sodium loss due to the lack of waste product wash. The sodium loss in a concentrated system without cake wash is probably greater than that which could be tolerated in a large scale FGD application. The loss is both costly and a potential water pollution problem.

Pilot Plant Testing - Data reported by FMC which were taken in a 3 week test, 8 hours per day, in the FMC mobile pilot plant processing a slip stream of industrial boiler flue gas are tabulated below:

Flue Gas

Boiler type: Stoker-fired coal boiler

Fuel: Coal, 4.8% sulfur

Flue gas flow: 2,578 ACFM @ 400°F

Flue gas composition: N₂ = 76.3%

O₂ = 7.6%

CO₂ = 11.4%

H₂O = 4.7%

100.0%

SO₂ inlet = 3363 ppm

SO₂ exit = 343 ppm

Particulate inlet = 2.4 gr/SCF

Absorption

SO₂ removal: 89+%

Absorber type: Dual throat venturi

L/G ratio: 13.6 gal./1000 ACF

Pressure drop: 10 in. H₂O

Liquor composition: pH 6-7

total sodium >2 Molar

active alkali >0.5 Molar

Regeneration

Lime stoichiometry: ≤ 105% based on SO₂ removed

Soda ash addition: ≤ 0.026 moles Na₂CO₃/mole SO₂
removed

Reactor residence time: 5 minutes

Thickener rise rate: 0.3-0.4 gpm/ft²

Solids Separation

Filter cake composition:	CaSO ₃ ·1/2 H ₂ O	47.93%
	Ash	13.95%
	CaCO ₃	1.14%
	Sodium as Na ₂ SO ₄	1.18%
	H ₂ O	35.80%
		100.00%

Solubles in waste product: 1.84% (dry basis)

Soluble sulfur loss: 0.027 moles soluble sulfur/
mole SO₂ removed

Filter cake wash: 2 displacement wash or less,
90% reduction in sodium

Discussion - The results reported are typical for operation of the FMC concentrated double alkali system in a high sulfur coal application. As previously discussed, the concentrated systems are better suited to high sulfur, low oxygen (rather than low sulfur, high oxygen) applications due to the necessity to control sulfate formed by oxidation in the system. Since sulfate formation rate in the system is related to the rate of oxygen absorption, sulfate will comprise a smaller fraction of the total soluble sulfur species in a concentrated system than in a dilute system. Thus the soluble sulfur loss is a small fraction of the SO₂ removed in these applications; however, it can be expected to be a larger fraction in low sulfur fuel applications if all other factors remain constant.

The FMC system is not designed for sulfate removal by any of the chemical mechanisms described under the term "sulfate removal"; thus, there must be a loss of soluble sulfate from the system at the rate at which it is formed. This soluble sulfur loss occurs as solubles contained in the waste product cake moisture. Apparently, operation at the very high ionic strengths employed by FMC minimizes the removal of soluble sulfate via the mixed crystal or solid solution mechanism described by equation (17). On the other hand, the high ionic strength liquor employed by FMC probably absorbs oxygen at a lower rate than the less concentrated systems, and thus reduces sulfate formation by reducing oxidation rate. The soluble solids level reported in the above data by FMC appears to be as low percentagewise as any of the values reported for other throwaway product systems in the U.S. In general, FMC has stressed lowering sulfate formation rates rather than methods of sulfate removal.

Another important factor which influences oxidation rate is scrubber type. FMC has conducted tests with a packed tower, venturi, valve tray (Koch) scrubber and a "disc and donut" scrubber, and has correlated oxidation rate with scrubber type. As expected, the scrubbers with the larger liquor residence times exhibit the higher oxidation rates.

Table 2 presents a summary of the FMC testing with the 3500 ACFM mobile pilot plant. A wide range of testing experience is shown for the mobile pilot plant. No scaling problems or other major reliability problems have been encountered in this testing.

c. Future Applications - FMC has four double alkali systems under construction, and a fifth in the design phase in the U.S. The largest planned U.S. application of the FMC system is a 150 Mw system for the control of SO_2 from the pulverized coal boilers at FMC's Green River, Wyoming, plant. The system is scheduled to start up in the summer of 1975. In addition, FMC has licensed the Ataka Construction and Engineering Company of Osaka, Japan, to install their system in Japan. The first application of the FMC technology in Japan will be for the control of SO_2 from a 45,000 lb/hr steam boiler at a paper manufacturing company. That installation is scheduled to start up in the first quarter of 1975.

A summary of the FMC systems planned for near future operation is given in Table 3.

Table 2. FMC 3500 ACFM MOBILE PILOT PLANT TESTING

Application	Test Time (Months)	SO ₂ Removal (%)	Particulate Removal (%)	Scrubber Type	Pressure Drop (in. H ₂ O)	Temp. (°F)	Flue Gas O ₂ Content (%)	Inlet SO ₂ (ppm)
1. SO ₂ and flyash, stoker boiler (FMC Plant So. Charleston, West Virginia)	6	90+	99+	Dual Throat Venturi	10-15	400	12	400-1400
2. SO ₂ and flyash stoker boiler	3	90+	99+	Dual Throat Venturi	10-15	350	10	1800
3. SO ₂ and flyash, stoker boiler (General Motors Plant)	4	90	99	Dual Throat Venturi	10	390	9-10	600-1200
4. SO ₂ and flyash on bark oil and coal fired boiler	2	90	99.5	Dual Throat Venturi	10	400	7	1000
5. SO ₂ and flyash, pulverized coal boiler	1	89+	99+	Dual Throat Venturi	10	400	7	3300
6. SO ₂ and flyash, pulverized coal electric power boiler	4	90+		Dual Throat Venturi, Koch Tray, Disc & Donut	4.5-10	250	5	1000-1200
7. SO ₂ control, sulfuric acid plant	1	90+		Dual Throat Venturi, Disc & Donut	4.5-10	120	6	1200

Table 3. FUTURE FMC DOUBLE ALKALI APPLICATIONS

<u>Application</u>	<u>Size (Mw, Equivalent)</u>	<u>Scrubber Type</u>	<u>SO₂ Inlet (ppm)</u>	<u>Start-Up Date</u>
1. Stoker fired industrial boiler system	50	Dual Throat Venturi	1800	First Quarter 1975
2. Oil and coal fired boiler (Firestone, Pottstown, Pa.)	3	Dual Throat Venturi	1400	December 1974
3. Pulverized coal boiler (FMC's Green River, Wyo. Plant)	150	Disc & Donut	1000	Mid 1975
4. Oil fired boilers	13	Disc & Donut	250	Construction to start in 1975
5. Heating plant coal boiler	14	Disc & Donut	---	Late 1974
6. Paper plant, boiler (Japan)	5	-	---	First Quarter 1975

General Motors

General Motors (GM) Corporation developed a dilute double alkali system through pilot testing in a 2800 CFM cross-flow packed bed scrubber at the GM Chevrolet, Parma, Ohio, plant. Flue gas for pilot testing was an isokinetic sampling representing 10% of the total flue gas flow from a boiler having a steaming capacity of 80,000 lb/hr and burning 2% sulfur coal with 100% excess air.

In the pilot plant, it was found that a 1 molar sodium solution would give reasonably good SO_2 absorption, while also regenerating caustic and precipitating sulfate as gypsum. It was found that a maximum concentration of 0.1 M hydroxide could be regenerated in a 1 M sodium solution. Increasing the sodium ion concentration above 1 M did not give appreciable increase in regenerated hydroxide concentration. Lowering the sodium ion concentration, however, gave a decrease in regenerated hydroxide concentration.

Optimum lime utilization was found to occur with high speed mixing, using near stoichiometric quantities of lime. Eighty percent conversion of lime was attained in 5 minutes.

Early in the pilot plant program, GM ran into the calcium scaling problem; but later, using soda ash softening, they were able to alleviate it. Reportedly, using soda ash for sodium make-up, they were able to reduce the calcium content in the scrubber loop from 400 to 250 ppm.

The pilot plant was operated for a period of about 4 months in Parma. Other pilot tests were conducted concurrent with the Parma tests at another General Motors plant in Livonia, Michigan.⁸

Based on the results of testing in the pilot units, General Motors designed and constructed a full scale system for SO_2 and particulate control at the Parma plant.

a. System Description⁵ - The full scale double alkali system installed at Parma is shown schematically (only one scrubber is shown) in Figure 5. The boiler system consists of four stoker fired boilers, two with steaming capacities of 100,000 lb/hr and two with 60,000 lb/hr, totalling 320,000 lb/hr or approximately equivalent to 32 Mw electric generating capacity. All of the boilers are equipped with mechanical dust collectors.

The double alkali system was designed to handle a steaming capacity of 400,000 lb/hr anticipating the addition of a fifth boiler in the future.

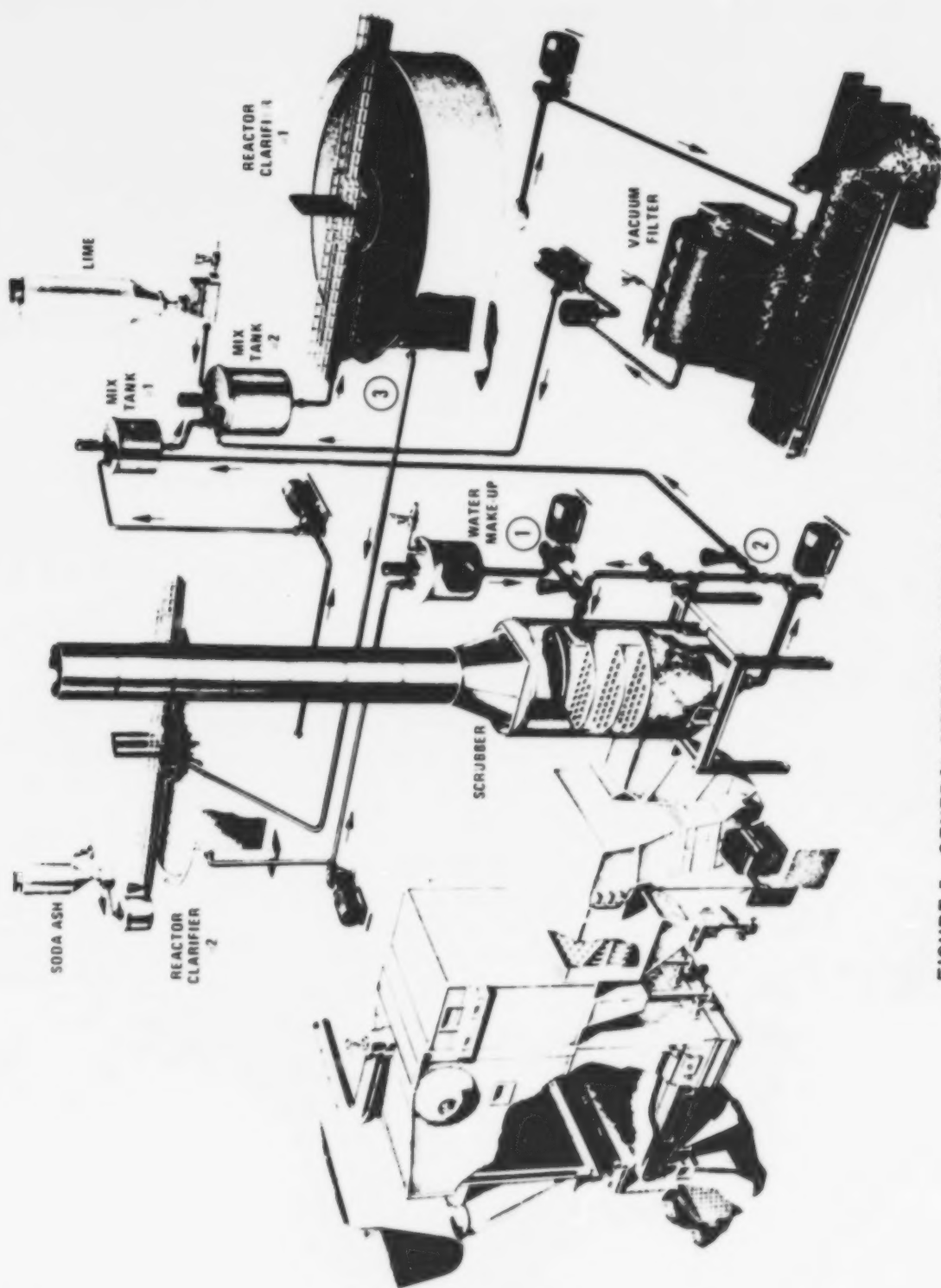


FIGURE 5. GENERAL MOTORS' DOUBLE-ALKALI SYSTEM.

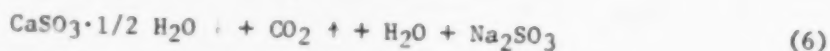
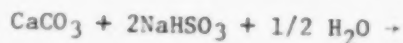
The absorption system consists of four Koch valve tray scrubbers containing three trays per scrubber. The trays have a type of floating bubble cap which rises and falls adjusting to gas flow. The upper portion of the scrubbers contain mesh mist eliminators. The scrubbers are constructed of 316-L stainless steel.

The regeneration system handles the bleed flow of spent liquor from the four scrubbers. It consists of two mix tanks or reaction tanks (CFSTR's), two slaker-feeders (one on standby) and two reactor clarifiers. One of two continuous drum filters makes the final solids separation (the second is on standby). The filters are equipped with spray nozzles to allow for cake wash.

The system operates similarly to the Envirotech dilute scheme, with the primary exception that there is no augmental oxidation; however, due to the fact that the boilers operate with high excess air, high oxidation rates in the system are expected, and have been experienced.

The primary functions of some of the major equipment are given below:

Mix Tank No. 1 - Spent scrubber solution containing Na_2SO_4 , Na_2SO_3 and NaHSO_3 is mixed with the slurry underflow from reactor clarifier No. 2 which should contain NaOH , Na_2SO_3 and CaCO_3 . This tank was designed to increase lime utilization by making use of the CaCO_3 that results from the softening reaction in reactor clarifier No. 2 (discussed later), to neutralize some of the bisulfite in the spent scrubber solution. The extent of CaCO_3 utilization, however, is highly dependent on the scrubber bleed stream pH. If this pH is not low enough, very little, if any, CaCO_3 will be utilized as shown in the following reaction (previously shown under limestone regeneration):



Mix Tank No. 2 - The primary reactions between lime and sulfite, bisulfite and sulfate begin to occur in this tank, precipitating gypsum and calcium sulfite. Effluent from mix tank No. 1 and filtrate from the filter are mixed with lime in this tank.

Reactor Clarifier No. 1 - Slurry feed is received from Mix Tank No. 2. Here the reaction between lime, sulfite, bisulfite and sulfate is continued. The overflow consists of clear regenerated liquor, and the underflow is a thickened slurry of waste solids which is fed to the filter for final solids separation.

b. Operating Experience - General conditions of operation and performance data for the full-scale system at Parma are tabulated below:

Source: four coal stoker fired boilers
 Fuel: coal, 2.3-2.6% sulfur, 2.5% sulfur average
 Boiler excess air: 50% large boilers
 120% small boilers
 Oxygen content: 8.5-11.5%
 Flow rate: 60,000 ACFM @ 350°F - Large boilers
 50,000 ACFM @ 550°F - Small boilers } design
 26,000-28,000 SCFM/scrubber, typical of
 recent operation
 (Normally, 2 of 4 boilers and scrubbers
 in operation)
 SO₂ inlet concentration: 900-1600 ppm (1200-1300 ppm
 average)
 Particulate inlet concentration: 0.3 gr/SCF

Absorber type: Koch valve tray scrubber, 3 trays
SO₂ removal: 88-98%
SO₂ exit concentration: 20-200 ppm
Particulate exit concentration: 0.05 gr/SCF (design value)
L/G ratio: 20 gal./1000 CF
Scrubber pressure drop: 7-1/2 in. H₂O
Scrubber liquor pH: inlet ~ 9.0
 outlet 5.5-6.0

Lime stoichiometry: 140-165% based on SO₂ removal
Regeneration liquor composition:

495

Solids Separation

Solid Waste composition:	Ca(OH) ₂	10-20%	dry basis
	Flyash	1-2%	dry basis
	Solubles (Na ₂ SO ₄)	4-5%	dry basis
	CaSO _x	Remainder	
	Moisture	50%	
Solid product wash:	Crude wash, 20% reduction of solubles		

Reliability - The Parma full-scale system was started up in March 1974. There were at least two scheduled shutdowns soon after start-up for inspection. After the second shutdown a problem developed which kept the system off line for about 3 weeks. As of early October 1974 the system has been in operation since May 23, without a total system shutdown. Since the total system consists of four scrubbers and four boilers, and the steam requirements have been such that usually only two boilers and scrubbers have been on line at any one time, it does not necessarily follow that each boiler/scrubber system was without problems for the whole period between May and October 1974. In fact, if and when a problem developed in one boiler/scrubber system, that system was shut down and another started up. This has been possible since the plant steam requirements have been well below the system capacity.

Of the problems which have been experienced since start-up, only one seems to be recurrent and is cause for individual scrubber shutdown: the top tray in the scrubbers has a tendency to scale with calcium carbonate. The other problems were relatively minor, and were possibly a result of lack of operating experience. The scaling problem is probably due to the fact that high pH liquor (pH above 12) from the regeneration system enters the top tray of the scrubber directly, causing local high pH regions on the top tray. The localized high pH regions absorb CO₂ from the flue gas, precipitating CaCO₃ scale. A solution for this problem is planned. Regenerated liquor will be mixed with scrubber recycle liquor (pH 5.5-6.0) so that pH on the top tray is controlled at a level low enough to prevent precipitation of CaCO₃ (equations (23) and (24)). At pH's below about 9, CaCO₃ is solubilized due to the carbonate/bicarbonate equilibrium:



A more complete discussion of the GM system and the operating problems experienced to date is given by Dingo and Piasecki.¹⁶

c. Discussion - As of early October 1974, the GM system at Parma has not been in steady state operation in the strict sense of the term. Some improvement and optimization of the system, including solution of the carbonate scaling problem, are still intended. The sulfate level in the system presently at 0.35M concentration is lower than the design level of 0.5M. Fluctuations in the sulfate level are indicative of poor control of soda ash make-up to the system. The high lime stoichiometry, 140-165%, is indicative of poor lime utilization. Closer control of the system should bring this lime utilization more in line with the lower lime stoichiometries in use in the other double alkali systems.

General Motors has entered into an agreement with EPA which will allow Arthur D. Little, Inc., as a contractor for EPA's Control Systems Laboratory, to conduct a test program at the Parma plant to more fully characterize the double alkali system. Through careful material balances in the system, calculated from flow rate measurement and chemical analytical data, a number of important parameters will be studied. Among these are included:

- SO₂ removal
- Process reliability (scaling phenomena)
- Sulfate control
- Waste characteristics
- Degree of closed loop operation
- Utilization of lime and other chemicals
- Oxidation rate

Zurn Industries

Zurn has designed, constructed, and started up a dilute double alkali system for control of SO_2 from two coal fired industrial boilers having a total steaming capacity (180,000 lbs/hour) equivalent to about 20 Mw electric generating capacity. The system was started up in early October 1974; consequently, no operating data can be reported at this time. The boilers burn coal containing about 3.0% sulfur. Generally, the system is patterned after the General Motors system in Parma. The scrubber system consists of two Zurn "Dustractor" scrubbers, containing 24 and 28 tubes, respectively. Briefly described, this is a multitube impingement/entrainment type scrubber capable of achieving very high internal L/G ratios. The system employs two series lime reactors and a sodium carbonate softening step. Provisions have been made for waste product washing on the rotary vacuum filters, to minimize sodium losses and associated potential secondary pollution problems.

A schematic of the Zurn double alkali system is shown in Figure 6.

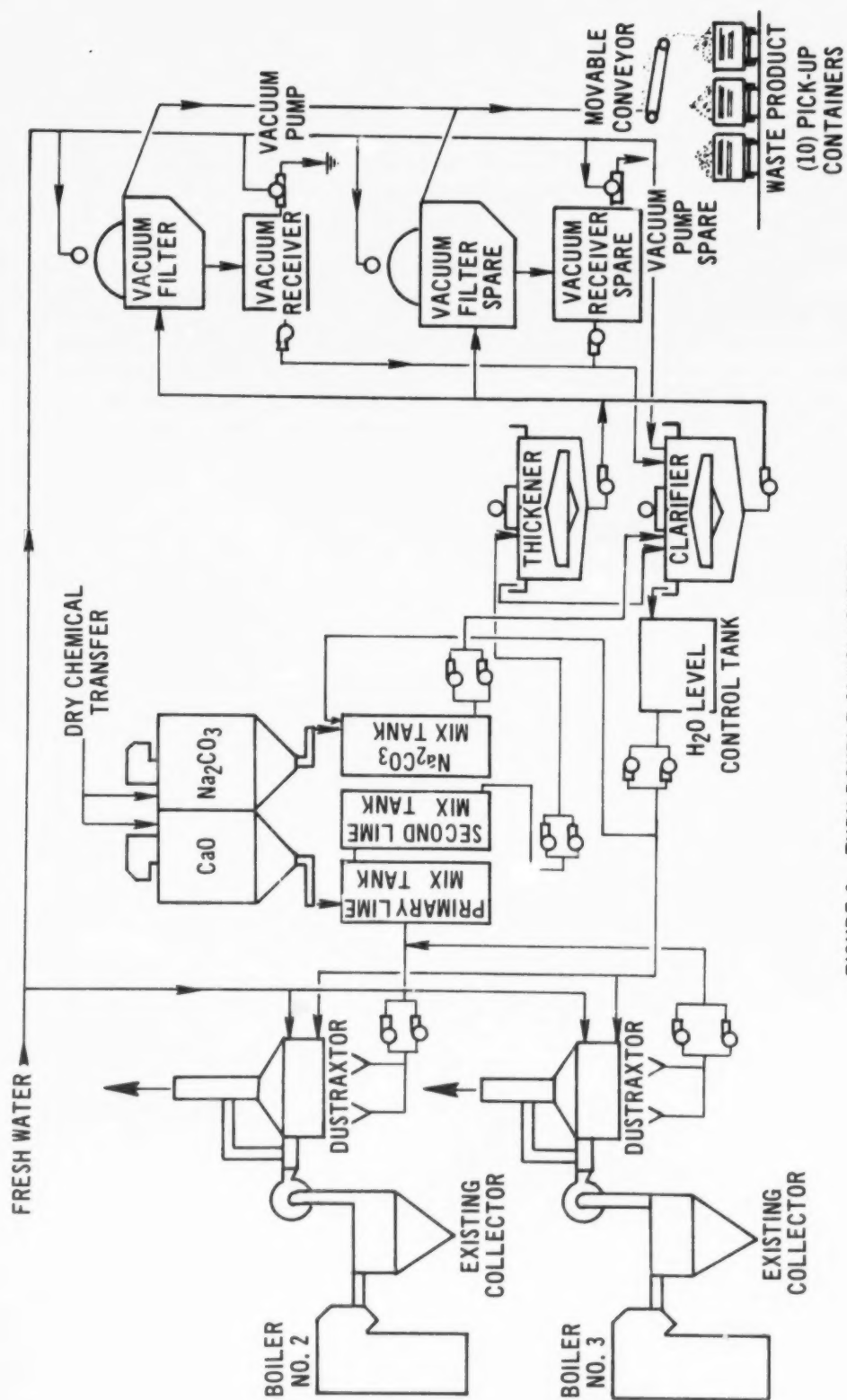


FIGURE 6. ZURN DOUBLE ALKALI SYSTEM

A. D. Little/Combustion Equipment Associates

Pilot testing by ADL/CEA in a 2000 ACFM pilot plant in the ADL complex in Cambridge, Massachusetts, in early 1973 led to the development of a concentrated double alkali system. Based on the results of that testing, CEA contracted with Southern Services, Inc., a subsidiary of The Southern Company (a Southeastern Utility Combine) to install a 20 Mw prototype test unit at the Scholz station of Gulf Power Company near Chatahoochee, Florida.

In May 1973, EPA contracted with ADL to conduct a laboratory and pilot plant study of various double alkali schemes of operation. About a year later, when construction of the prototype system was well underway, the EPA contract with ADL was expanded to include prototype testing of the ADL/CEA system at Gulf Power. A good description of the ADL/CEA pilot plant and the Gulf Power 20 Mw prototype system is given in the paper entitled, EPA-ADL Dual Alkali Program--Interim Results.³ Also contained therein are operating conditions and test results obtained under the EPA program to date. The system developed by ADL/CEA is among the various schemes tested under the program.

Kawasaki/Kureha

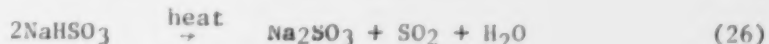
Kureha Chemical Industry Co., Ltd., developed a concentrated limestone double alkali process in a small pilot plant. Initial testing led to the construction of a larger pilot plant (3000 SCFM) jointly with Kawasaki Heavy Industries, Ltd., in July 1972. Pilot testing was completed in March 1973. Since then, these two companies have jointly developed a commercial double alkali process which is in operation at the Shinsendai station of Tohoku Electric Power. This unit processes one quarter of the flue gas from a 600 Mw oil-fired boiler (150 Mw equivalent).

a. System Description - A schematic diagram of the system installed at Tohoku Electric is shown in Figure 7. The system consists of an SO₂ absorption section, a limestone regeneration section, a sulfate removal section and a gypsum production section.

Absorption - Flue gas from the oil fired boiler passes through an electrostatic precipitator before entering the system. The gas is moved by a hot fan, forced draft with respect to the scrubber system. There is a precooler section at the entrance to the scrubber in which the hot incoming gas is adiabatically cooled with recycle liquor. The scrubber is a grid-packed type consisting of two separate stages of packing; gas flows upward through the packing. Scrubber recycle liquor and effluent from the top stage are fed to the top of the bottom stage. The bottom stage operates at an L/G ratio of about 12 gal./1000 CF. Regenerated liquor, only, is fed to the top stage thus maximizing the pH in this stage and achieving good countercurrent mass transfer resulting in high SO₂ removal efficiency. The total SO₂ removal in the system is split, 90% in the bottom stage and 10% in the top stage (this does not imply 100% removal of the incoming SO₂). Flue leaving the absorber is reheated to 285°F by direct firing of 1.2% sulfur oil.

Regeneration - The regeneration reaction consists of neutralization of NaHSO₃ with finely ground limestone (equation (6)). The limestone is ground to 10 microns in a wet vertical tower mill. Scrubber effluent liquor bleed taken from the bottom stage recycle stream is filtered to remove oil soot before it is pumped to the regeneration reactors to be reacted with limestone. The reactor system consists of five closed reactors in series which are operated at about 90°C to promote the reaction between limestone and NaHSO₃ in the scrubber bleed. The solid product produced in the reactors is separated from the mother liquor and washed in a vacuum filter. The liquor is recycled to the scrubber system, and the solids, mostly CaSO₃·1/2 H₂O are fed to the oxidizer system and to the sulfate removal section.

Sulfate Removal - This system is designed to remove sulfate formed in the scrubber as gypsum via reaction with sulfuric acid and calcium sulfite according to reaction (16). The process is fed with a bleed stream of scrubber effluent, a portion of the calcium sulfite cake (washed) from the filter and sulfuric acid. In order to maximize sulfuric acid efficiency, a small absorber and stripper are incorporated in this sulfate treatment loop. The incoming scrubber bleed is acidified in the absorber with SO₂. SO₂ is supplied by the stripper which converts NaHSO₃ to Na₂SO₃, stripping off SO₂ via thermal decomposition according to the following reaction:

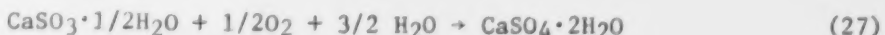


This reaction is the thermal regeneration reaction used in the Wellman-Lord process which is now commercialized by Davy Powergas. Feed liquor to the stripper is the acidic mother liquor from the sulfuric acid reactor in which the sulfate removal occurs. Stripped liquor from the stripper is recycled to the regeneration reactor system.

Feed to the sulfuric acid reactor is thus already at a low pH, and therefore requires less sulfuric acid addition to achieve a given conversion of Na₂SO₄ to gypsum.

Effluent slurry from the sulfuric acid reaction is centrifuged to separate the gypsum produced from acidic mother liquor. The separated solids are transferred to the oxidizer system together with the calcium sulfite solids produced in the regeneration reaction. The centrifugate, as previously stated, is recycled to the stripper to allow recovery of acid values.

Oxidation - The oxidizer developed by Kawasaki/Kureha is considered to be proprietary. Oxidation air is supplied at 5 psig. Repulped calcium sulfite slurry from regeneration and gypsum from the sulfate removal step are pumped to the oxidizer. Sulfuric acid is added to adjust pH in the reaction to attain the desired conditions for the oxidation reaction. Calcium sulfite is oxidized to gypsum by reaction with atmospheric oxygen according to the overall reaction:



The reaction mechanism may be acidic dissolution of CaSO₃·1/2H₂O followed by oxidation of the bisulfite to sulfate regenerating the acid value to dissolve additional CaSO₃·1/2H₂O. Gypsum product is separated from oxidizer effluent and washed in centrifuges. Spent off-gas from the oxidizer is recycled through the scrubber to remove SO₂ which was emitted to the acidic oxidation reaction.

b. Operating Experience - Typical conditions and performance data for the Tohoku Electric FGD system are tabulated below:

Flue Gas

Source: oil fired boiler
Fuel: 1.2-1.5% sulfur oil
Oxygen content: 3%
Flow rate: 247,000 SCFM (420,000 Nm³/hr)
SO₂ inlet concentration: 600-800 ppm
Particulate inlet: 0.0009-0.0044 gr/SCF (2-10 mg/m³)
Reheat: direct fire with 1.2% sulfur oil to 285°F
(140°C)

Absorber System

Absorber type: grid packed, 2 packing stages
SO₂ removal: 98%
SO₂ exit concentration: <10 ppm
L/G ratio: 12 gal./1000 CF (1 kg/kg), bottom stage
Scrubber pressure drop: 7.9 in. H₂O (200 mm H₂O)
Scrubber liquor pH: 6.9-7.6 inlet to top stage
6.3-6.5 bottom stage recycle
Recycle liquor composition: 5-10% Na₂SO₄
5-8% NaHSO₄
6-9% Na₂SO₃
Oxidation rate: 5%

Regeneration

Limestone stoichiometry: 100% of SO₂ inlet
Sodium make-up rate: 0.018 moles Na/mole SO₂ absorbed
Reactor system residence time: 3 hrs (total, 5 reactors)
Reactor % solids: 5% (essentially all CaSO₃·1/2H₂O)
Reactor filter cake solids: 40%
Regenerated liquor: 6.9-7.6 pH
20% total dissolved solids
<10 ppm Ca
Limestone particle size: 10 μ

Solids Production (Gypsum)

Water content: 5%
Sodium content: <300 ppm on dry basis
Particle size: 30x200 μ

Reliability - The full-scale unit started up in January/February 1974. Based on reports as of August 1974, there have been no significant reliability problems. The only problem noted during a visit in Japan by a representative of EPA was the plugging of the oil soot filter.

Discussion - The system is said to be a closed loop system in that there is no intentional water purge; however, based on the rate of water addition to the system, and a comparison of the rate of sodium make-up with the sodium content of the product, it appears that there is some sodium loss from the system other than the amount in the product gypsum. The overall system is fairly complex when compared to some of the systems under development in the U.S. The reasons for the additional complexity appear to be related to a) the system produces a relatively pure saleable product, b) it is designed to achieve extremely high SO₂ and particulate removal, and c) limestone rather than lime is used as the source of calcium. Apparently, there has been considerable effort devoted to minimization of reagent usage; i.e., sulfuric acid, limestone and sodium.

The gypsum product has been used in cement manufacture; however, it also meets the more stringent specs for wallboard production.

The very low oxidation (5%) in the absorption system is attributed by Kawasaki/Kureha to proprietary know-how.

c. Plans for Future Testing and Full-Scale Application - Kawasaki/Kureha plans to pilot-test an electrolytic process step which was previously piloted by Yuasa/Ionics in Japan. This process step is being evaluated to replace the sulfuric acid reactor/absorber/stripper operation for sulfate removal. The electrolytic sulfate removal scheme is similar to that used in the Stone & Webster/Ionics process, equation (18). Testing was scheduled for October 1974, in a 3-5 Mw pilot plant consisting of two 5 ft² cells capable of decomposing 10 lbs/hour of Na₂SO₄. Use of the electrolytic step is expected to reduce sodium make-up requirement to the system by 50%, and is anticipated to reduce capital costs of the overall system.

A number of additional large scale utility double alkali systems are presently being engineered and/or constructed by Kawasaki/Kureha. These are tabulated below:

<u>Utility</u>	<u>Capacity (Mw)</u>	<u>Start-Up</u>
Sikoku Electric Power	450	August 1975
Sikoku Electric Power	450	October 1975
Tohoku Electric Power	350	October 1976
Kyushu Electric Power	250	May 1977
Kyushu Electric Power	250	May 1977

Showa Denko KK/Ebara

Showa Denko KK and Ebara Manufacturing Co., Ltd., of Japan jointly developed a concentrated double alkali process using limestone as the calcium source, and producing a saleable gypsum by-product. Development began in a 5,900 SCFM pilot plant at the Kawasaki plant of Showa Denko in 1971. This has culminated in the construction and operation of a 150 Mw system on an oil fired boiler at Showa Denko's plant in Chiba.

a. System Description - Figure 8 is a schematic representation of the Showa Denko/Ebara process. Basically it is similar to the Kawasaki/Kureha system. The full scale unit (150 Mw) uses four inverted "vertical cone" scrubbers (similar to the Zurn or Bahco type). Liquor is entrained at the bottom of a conical draft tube (truncated apex) by the entering flue gas and the mixture passes up the tube. At the top of the tube (conical base) there is a disengaging section where liquor separates and falls back to the bottom of the scrubber.

In the reactor system, limestone is reacted with spent scrubber effluent to neutralize the NaHSO_3 producing Na_2SO_3 for recycle to the scrubber and $\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$ (equation (16)) which is sent to an air oxidizer to be converted to gypsum by air oxidation (equation (28)). A slip stream of spent scrubber liquor is taken to a sulfate treatment section which converts Na_2SO_4 to gypsum and NaHSO_3 by reacting it with sulfuric acid and calcium sulfite according to reaction (16). Gypsum from this reaction containing some unconverted calcium sulfite is combined with the calcium sulfite produced in the limestone neutralization reaction and pumped to the air oxidizer system. Sulfuric acid is added to adjust the pH in the oxidizer system to achieve optimum conditions for the calcium sulfite oxidation.

The system operates in an open loop since the amount of water required to wash the gypsum product to below 300 ppm sodium content is greater than the system water losses through evaporation, water of hydration, and moisture in the final product. The sodium content specification for wallboard manufacture is less than 300 ppm. The corresponding specifications for gypsum used in cement manufacture is 900 ppm.

b. Operating Experience - General conditions and performance data for the full scale system are tabulated below:

Flue Gas

Source: oil fired 150 Mw boiler
Fuel: 2.5-3.0% sulfur
Particulate inlet concentration: 0.8 gr/scf
SO₂ inlet concentration: 1200-1500 ppm
Reheat: direct fired with 1% sulfur oil to 285-300°F

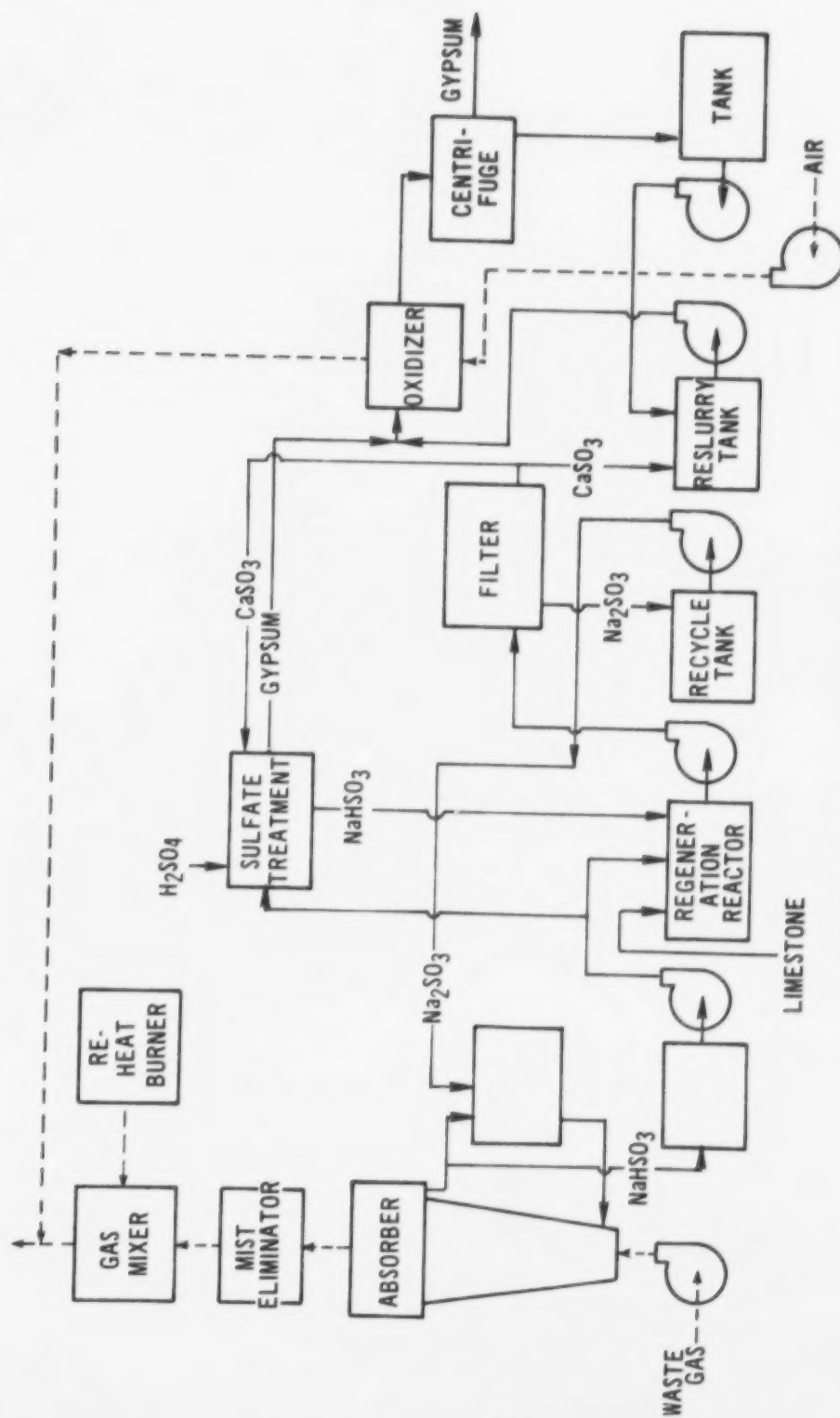


FIGURE 8. SHOWA DENKO SODIUM-LIMESTONE PROCESS

Absorber System

Absorber type: 4 vertical cone scrubbers
SO₂ exit concentration: 60-100 ppm after reheat
with 1% sulfur oil
Particulate exit concentration: 0.025-0.035 gr/SCF
L/G ratio: 7.5 gal./1000 CF
Scrubber liquor pH: 6.3 exit
Oxidation rate: <10%

Regeneration Chemical Usage

Limestone utilization: 96%
Sulfuric acid usage: 0.2 moles H₂SO₄/mole SO₂ absorbed
Sodium usage: 0.10-0.12 moles of NaOH/mole SO₂ absorbed

Oxidizer

Suspended solids content: 15%

Gypsum By-Product

Na content: <300 ppm
Water content: 7-8%

Reliability - The full scale system was started up in June 1973. It was shut down for a period in March 1974 for inspection. As of March 1974, the scrubbers were said to have been near 100% reliable, and they were found to be clean in the March inspection. The regeneration system had given some trouble but since there is adequate solution surge capacity, regeneration difficulties have not affected boiler reliability. The trouble has been mainly pipe plugging which was not considered to be chemical scaling.

SUMMARY AND CONCLUSIONS

Based on the initial performance and reliability demonstrated in various double alkali pilot plants, and in an increasing number of full scale applications in the U.S. and Japan, it appears that this technology will soon become a viable alternative throwaway process to the lime/limestone processes which are now the prevalent FGD processes used with fossil fueled utility boilers. The total operating and planned full-scale and prototype utility and industrial applications of double alkali technology now number 17 with individual applications with total gas flow control equivalent to 2375 Mw of electric generating capacity. Summaries of significant pilot testing, and operating and planned full scale systems are given in Tables 4 and 5, respectively.

Performance of double alkali systems with respect to SO₂ removal is well established. Over 99% SO₂ removal has been achieved with lower than 10 ppm SO₂ concentration in the treated flue gas.

Potential problems with waste disposal from these systems are present due to the presence of a certain amount of soluble sodium salts which could cause water pollution problems in the surface and ground water in the vicinity of disposal sites. A number of techniques to reduce the soluble sodium salts in the waste product have been tested; however, it appears that there will inevitably be a higher level of soluble salts in waste product from double alkali systems than from lime/limestone systems. With present U.S. technology it appears that the minimum incremental amount of solubles in a double alkali waste product (over that typically present in lime/limestone waste product) is about 1-2% on a dry solids basis. Development of sludge fixation technology to change the wet solid waste product to a hard unleachable solid could conceivably reduce this potential problem.

Actual costs for full scale utility boiler applications are not available. Estimates for these applications based on the best available information put double alkali system capital costs in the range of lime/limestone system costs; however, these types of estimates are necessarily subject to qualification.

In the U.S., development has stressed the use of lime rather than limestone as the calcium source in all of the full scale industrial applications and in most of the pilot plant testing in both dilute and concentrated systems. In Japan, limestone rather than lime is used for concentrated systems in full scale utility boiler applications apparently for the operating cost benefit which is derived through the use of the less expensive regenerant. Also, due to differences in the economies of the two countries and state-of-the-art of technology, production of by-product gypsum from the Japanese double alkali systems is prevalent;

Table 4. SUMMARY OF SIGNIFICANT PILOT PLANT DOUBLE ALKALI TESTING

<u>System Developer</u>	<u>Pilot Plant Location</u>	<u>Size (Gas rate, CFM)</u>	<u>Active Alkali</u>	<u>Calcium Source</u>
Envirotech	Utah Power & Light Company electric power boiler, coal fired	2500	Dilute, Conc.	Lime
FMC	At least 7 different locations, coal and oil fired industrial boilers, elec. power boilers, acid plant	3500	Conc.	Lime
General Motors	GM Parma Plant GM Livonia Plant coal fired industrial boiler	2800	Dilute	Lime
A. D. Little/ Combustion Equipment Associates/EPA	ADL Facility Cambridge, Massachusetts Natural gas furnace	2000	Dilute, Conc.	Lime, Limestone
Kawasaki/Kureha	Oil fired boiler, Japan	3000	Conc.	Limestone
Showa Denko KK	Oil fired boiler, Japan	5900	Conc.	Limestone

Table 5. SUMMARY OF OPERATING AND PLANNED FULL SCALE DOUBLE ALKALI SYSTEMS

<u>System Operator</u>	<u>System Application</u>	<u>Vendor or Developer</u>	<u>Size (%, Equivalent)</u>	<u>Active Alkali</u>	<u>Calcium Sources</u>	<u>Start-up Date</u>
FMC						
Modesto Calif. Plant	Reduction kilns	FMC	10 (Gas Rate)	Conc.	Lime	Dec. 1971
Shouwa Denko KK Kawasaki Plant, Japan	Oil-fired elec. power boiler	Shouwa Denko/Ebara	150	Conc.	Limestone	June 1975
Tohoku Electric Shinsendai Sta., Japan	Oil-fired utility boiler	Kawasaki/Kureha	150	Conc.	Limestone	Jan. 1974
General Motors Parma, Ohio Plant	4 coal fired industrial boilers	General Motors	40 (Regen) 32 (Gas Rate)	Dilute	Lime	Mar. 1974
(Unspecified) U.S.	2 coal fired industrial boilers	Zorn Industries	20	Dilute	Lime	Oct. 1974
Gulf Power Company Scholz Plant (Southern Services)	Coal fired utility boiler-prototype	A.D.Little/ Combust. Equip. Associates	20	Conc.	Lime	(Nov. 1974)
Firestone Pottstown, Pa.	Coal & oil fired industrial boiler-demonstration	FMC	3	Conc.	Lime	(Dec. 1974)
(Unspecified) U.S.	Coal fired industrial heating plant	FMC	14	Conc.	Lime	(Dec. 1974)
(Unspecified) U.S.	Coal fired industrial boiler system	FMC	50	Conc.	Lime	(early 1975)
(Unspecified) Japanese Paper Plant	Oil fired industrial boiler	FMC/Ataka	5	Conc.	Lime	(early 1975)

Table 5 (Continued). SUMMARY OF OPERATING AND PLANNED FULL SCALE DOUBLE ALKALI SYSTEMS

<u>System Operator</u>	<u>System Application</u>	<u>Vendor or Developer</u>	<u>Size (Mw, Equivalent)</u>	<u>Active Alkali</u>	<u>Calcium Sources</u>	<u>Start-Up Date^a</u>
FMC Green River, Wyo. Plant	Coal fired electric power boiler	FMC	150	Conc.	Lime	(mid 1975)
Sikoku Electric Power, Japan	2 Oil fired utility boilers	Kawasaki/ Kureha	900 (2-450 Mw)	Conc.	Limestone	(Aug.-Oct. 1975)
Tohoku Electric, Japan	Oil fired utility boiler	Kawasaki/ Kureha	350	Conc.	Limestone	(Oct. 1976)
(Unspecified) U.S.	Oil fired industrial boiler	FMC	13	Conc.	Lime	(Late 1976)
Kyushu Electric Power, Japan	2 Oil fired utility boilers	Kawasaki/ Kureha	500 (2-250 Mw)	Conc.	Limestone	(May 1977)

^a Dates in parentheses are projected start-up dates

whereas, production of a throwaway solid waste product is the general rule with development of the technology in the U.S.

The development of double alkali technology has obviously stemmed (and benefitted) from both lime/limestone and other soluble sodium scrubbing technology development. Possibly as a result of this and certain inherent advantages of soluble alkali scrubbing, it appears that the reliability established at this point in the development of double alkali technology is greater than that which had been established for the lime/limestone systems at a corresponding stage of development.

REFERENCES

1. Ponder, W. H., "Status of Flue Gas Desulfurization Technology For Power Plant Pollution Control." Presented at Thermal Power Conference, Washington State University, Pullman, Washington, October 4, 1974.
2. Epstein, M., R. Borgwardt, et al., "Preliminary Report of Test Results from the EPA Alkali Scrubbing Test Facilities at the TVA Shawnee Power Plant and at Research Triangle Park." Presented at Public Briefing, Research Triangle Park, North Carolina, December 19, 1973.
3. LaMantia, C., et al., "EPA-ADL Dual Alkali Program Interim Results." Presented at EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.
4. Johnstone, H. F., et al., "Recovery of Sulfur Dioxide from Waste Gases." Ind. & Eng. Chem., Vol. 30, No. 1, January 1938, pp 101-109.
5. Phillips, R. J., "Operating Experiences with a Commercial Dual-Alkali SO₂ Removal System." Presented at the 67th Annual Meeting of the Air Pollution Control Association, Denver, Colorado, June 9-13, 1974.
6. Cornell, C. F. and D. A. Dahlstrom, "Performance Results on a 2500 ACFM Double Alkali Plant for SO₂ Removal." Presented at the 66th Annual Meeting of A.I.Ch.E., Philadelphia, Pennsylvania, November 11-15, 1973. Condensed version of the paper appeared in December 1973 CEP.
7. Kaplan, N., "An EPA Overview of Sodium-Based Double Alkali Processes - Part II Status of Technology and Description of Attractive Schemes." Presented at the EPA Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973.
8. Phillips, R. J., "Sulfur Dioxide Emission Control for Industrial Power Plants." Presented at the Second International Lime/Limestone Wet-Scrubbing Symposium, New Orleans, Louisiana, November 8-12, 1971.
9. Brady, J. D., "Sulfur Dioxide Removal Using Soluble Sulfites." Presented at Rocky Mountain States Section Air Pollution Control Association, Colorado Springs, Colorado, April 30, 1974.

10. Cornell, C. F., "Liquid-Solids Separation in Air Pollutant Removal Systems." Presented at the ASCE Annual and National Environmental Engineering Convention, Kansas City, Missouri, October 21-25, 1974.
11. Selmeczi, J. G. and R. G. Knight, "Properties of Power Plant Waste Sludges." Presented at the Third International Ash Utilization Symposium, Pittsburgh, Pennsylvania, March 13-14, 1973.
12. Ellison, W., et al., "System Reliability and Environmental Impact of SO₂ Scrubbing Processes." Presented at Coal and The Environment, Technical Conference, Louisville, Kentucky, October 22-24, 1974.
13. Rochelle, G. T., "Economics of Flue Gas Desulfurization." Presented at EPA Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973.
14. "Sulfur Dioxide and Flyash Control", FMC Corporation Technical Bulletin. FMC Corporation, Air Pollution Control Operation, 751 Roosevelt Road, Suite 305, Glen Ellyn, Illinois 60137.
15. McGlamery, G. G. and R. L. Torstrick, "Cost Comparisons of Flue Gas Desulfurization Systems." Presented at the EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.
16. Dingo, T. and E. Piasecki, "General Motor's Operating Experience with a Full-Scale Double Alkali Process." Presented at the EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.

Initial Operating Experiences With A Dual-Alkali SO₂ Removal System

Part I

PROCESS PERFORMANCE WITH A COMMERCIAL
DUAL-ALKALI SO₂ REMOVAL SYSTEM

Thomas T. Dingo/General Motors Corporation

Part II

EQUIPMENT PERFORMANCE WITH A COMMERCIAL
DUAL-ALKALI SO₂ REMOVAL SYSTEM

Edmund J. Piasecki/General Motors Corporation

for presentation at

**ENVIRONMENTAL PROTECTION AGENCY
SYMPOSIUM ON FLUE GAS
DESULFURIZATION**

ATLANTA, GEORGIA
NOVEMBER 4-7, 1974

517

Preceding page blank

ABSTRACT

The first commercial demonstration of dual-alkali wet scrubbing went on-line in March, 1974, at the General Motors Chevrolet plant in Parma, Ohio. The SO₂ system was installed at a cost of \$3.5 million and was designed to handle a steaming capacity of 400,000 pounds of steam per hour. In this system, a dilute sodium hydroxide solution absorbs SO₂/particulate matter from the flue gas. The spent scrubbing solution is then recausticized with lime and reused in the scrubbers while precipitated sulfur compounds are dewatered and sent to landfill. The system is currently undergoing a shakedown and evaluation period.

This paper is an update of operating experiences. Equipment performance, SO₂ removal capabilities, process reliability and control, sulfate formation, chemical utilization, calcium ion control and waste cake characteristics are reviewed.

PART I

PROCESS PERFORMANCE WITH A
COMMERCIAL DUAL-ALKALI SO₂ REMOVAL SYSTEM

Thomas T. Dingo
General Motors Corporation
Manufacturing Development
Warren, Michigan

For Presentation at Environmental Protection
Agency Symposium on Flue Gas Desulfurization

Atlanta, Georgia
November 4-7, 1974

PROCESS PERFORMANCE WITH A
COMMERCIAL DUAL-ALKALI SO₂ REMOVAL SYSTEM

I. INTRODUCTION

General Motors, like everyone else, has been caught in the energy squeeze. Natural gas and fuel oil are not readily available as in the past. Thus, because of the long term availability of coal over oil or natural gas, the Corporation has made every effort to use this fuel wherever economically possible within existing pollution codes.

GM is already a large industrial user of coal in the generation of steam for process uses. In 1973, the Corporation consumed close to 2.0 million tons of coal. Strict air quality standards, however, in many urban areas have made the use of coal very difficult. In some cities, even switching to a low sulfur coal would not result in meeting air quality requirements. These situations have forced us to look to yet unproven sulfur dioxide emission control technology.

In 1968, in anticipation of more stringent pollution codes coupled with shortages and drastically high cost for low sulfur fuels, the Corporation initiated a program to determine if control technology could be developed for its industrial powerhouses. The criteria in developing such technology were as follows:

1. The system must afford high process and equipment reliability.
2. It must be simple and not dictate the operation of the powerhouse.
3. It must not be a chemical plant and must produce a by-product that can be easily disposed.
4. It must be economically competitive with other alternatives.

Scrubbing with a soluble salt coupled with regeneration and reuse of the scrubber effluent appeared to satisfy our criteria best. Pilot work was started in 1969 and the system developed has come to be known as the GM Double Alkali SO₂ Control System. The first commercial demonstration of this system is the \$3.5 million installation at GM's Chevrolet-Parma plant near Cleveland, Ohio, Figure 1. The system was placed on line in March, 1974 and has been going through shakedown and adjustment.

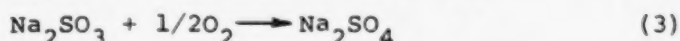
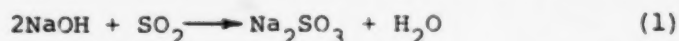
It is the purpose of this paper to describe the process chemistry and equipment and to relate the performance characteristics of the system to date.

II. PROCESS DESCRIPTION

The General Motors dual-alkali process involves the use of a dilute alkaline stream of sodium hydroxide (NaOH) to remove sulfur oxides from the boiler gases and alkaline lime (CaO) to regenerate the spent scrubbing liquor by precipitating the absorbed sulfur to reform an hydroxide solution. It is the purpose of this section to describe the process chemistry through the absorbing, regeneration, softening and neutralization sections of our system. Figure 2 shows a process schematic.

SCRUBBER CHEMISTRY

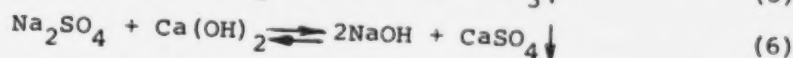
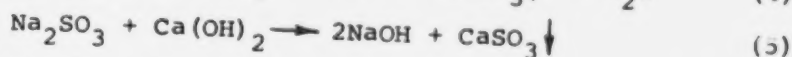
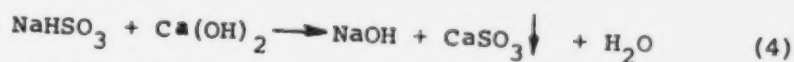
The reactions that take place in the absorbing section of our system are straight forward and are shown below:



Sodium hydroxide reacts with sulfur dioxide to form sodium sulfite (Reaction 1). The sodium sulfite, then can react further with sulfur dioxide and water to form sodium bi-sulfite (Reaction 2). In addition, some of the active sulfite is oxidized to sodium sulfate due to the presence of excess oxygen (Reaction 3).

REGENERATION CHEMISTRY

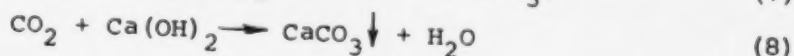
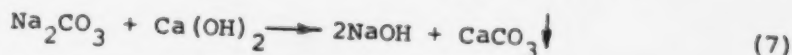
Because of the formation of Na_2SO_4 , the regeneration section has to be capable of regenerating both sulfate and sulfite. The sulfate reaction, however, is quite difficult because of the relative solubility of the product, calcium sulfate. In addition, the sodium sulfate cannot be causticized in the presence of high SO_3^{2-} or OH^- because Ca^{++} level are held below CaSO_4 solubility product. Thus, to provide effective sulfate regeneration, the system must be operated at a dilute OH^- concentration ($< 0.14\text{M}$) while maintaining sufficient levels of sulfate ($> 0.4\text{M}$) to affect calcium sulfate precipitation. In actual practice, we try to maintain our feed solution at 0.1 Molar OH^- and 0.5 Molar SO_4^{2-} . The chemical reaction that take place in the regeneration sections are:



SOFTENING CHEMISTRY

Because of the difficulty in regenerating sodium sulfate (Reaction 6), an excess of lime must be added. This causes dissolved calcium ion levels near 800ppm, which, if allowed to enter the scrubber, could precipitate as calcium salts and cause subsequent scaling.

To alleviate this problem, a softening step was added to our flowsheet. In this step, soda ash and carbon dioxide are added to precipitate the dissolved calcium as a carbonate. The reactions are:

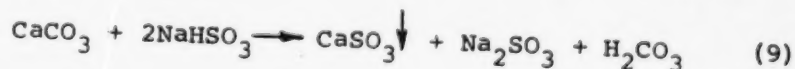


The resultant calcium levels now leaving the softening clarifier can be as low as 80ppm, but are controlled to approximately 300ppm.

A double media softening is used to maintain a chemical balance on the system. The main function of the soda ash is to replenish the sodium losses from the vacuum filter. Because of this, the soda ash make-up is not sufficient to lower the dissolved calcium to the desired level. The carbon dioxide, then, supplies the additional softening required. As shown in Reaction 8, however, the carbon dioxide lowers the free hydroxide available for scrubbing. From an economic and chemical utilization standpoint, the system should be operated with the minimum softening to avoid scrubber scaling.

NEUTRALIZATION CHEMISTRY

If the calcium carbonate formed through softening is not reused in the system, the lime utilization suffers drastically. In a sense, calcium would be added in one section and thrown out in the next. The only place where the calcium carbonate could be used back in the system is with the spent scrubbing liquor. The sodium bi-sulfite would react with the carbonate as shown in the following reaction:



Thus, a reaction tank to pretreat the scrubber effluent before lime addition was added to the system. Complete use of the calcium carbonate is dependent on the oxidation rate, low sulfate formation, in the scrubbers.

III. EQUIPMENT DESCRIPTION

The Chevrolet-Parma Plant was chosen as a demonstration site for our process because it is located near Cleveland where a rigid SO_2 emission limitation exists. This code requires us to burn an equivalent of 0.625% sulfur coal. The plant consumes approximately 55,000 tons/year of 2.0 to 3.0% + sulfur coal. The total steam generating capacity of the facility is 320,000 lbs/hr. from four spreader stoker fired boilers. In anticipation of an additional boiler, however, the sulfur dioxide control system was designed to handle a steam capacity of 400,000 lbs/hr. The equipment used in our double-alkali system is described below.

ABSORBERS

Four scrubbers are utilized in our system, one for each boiler. They are approximately 12 ft. in dia. and 20 ft. high, Figure 3. In these units, the boiler gases enter through a prequench-section at the bottom of the scrubber and flow upward through three absorption trays to exhaust. Scrubber feed and recycle are added at the top of the scrubber, flowing counter-current to the gases down to a recycle tank. Each of the absorption trays contains a series of movable bubble caps that adjust to changes in gas flows. Each unit is designed to give us a 3:1 turn down capacity. This type of absorber gives us good SO_2 removal at a relatively low recycle pH and good particulate removal. The particulate loading from our units must be near 0.05 grain per scf.

The scrubbers are constructed of 316-L stainless steel and are designed to operate at a total system pressure drop of 7.5 in. H_2O . They are also rated for a maximum liquid to gas ratio of 20:1 of which about 20% of the liquid is scrubber feed.

REACTORS

The bleed from the recycle tanks, Figure 4, is pumped through two back mix reactors in series. They are carbon steel in construction and are designed with a 6000 gallon capacity to give a 5 minute retention time at full flow. In the first reactor, a CaCO_3 slurry from the softener clarifier is added for neutralization. In the next reactor, lime slurry is added for recausticization. Both reactors are mounted on the side of the scrubber building as shown in Figure 5.

CLARIFIERS

From the reactors, the regenerated solution, laden with fly ash and calcium precipitates, flows by gravity to two 300,000 gal reactor clarifiers in series, Figure 6. Each clarifier is 60 ft. in dia. and has a maximum 0.46 gpm per square foot rise rate. In the primary clarifier, the solution goes through additional reaction, if needed, and solid separation. In the second clarifier, the solution is softened with Na_2CO_3 and CO_2 . A regenerated, clarified, and softened solution then overflows back to the scrubber building.

VACUUM FILTERS

The underflow from the primary clarifier (approximately a 15% slurry) is pumped to a sludge holding tank for batch vacuum filtering. Two 500 square foot filters are available, each with a maximum capacity of 50 tons of cake per day, Figure 7. The cake from the filters is dumped into trucks for disposal as a landfill material. The filtrate is returned to the system at the clarifiers.

CONTROL PANEL

To assure proper operation of the system, an elaborate panel was designed to simplify the control and monitoring functions, Figure 8. From here, the operator has positive indication of what is happening in the scrubbers, on the floor, and outside of the building and can change chemical addition rates as required.

The control theory of the double alkali system is relatively simple. First, caustic is fed to the scrubbers on the basis of the boiler steam load and trimmed by the pH in the recycle line. Steaming rate is a good indicator of coal usage and subsequent pounds of sulfur dioxide being formed. Since a certain pH range is desirable for proper SO_2 removal, pH control was also combined. Also, since the scrubber feed is proportional to the caustic being supplied to the scrubbers, this signal was selected as a basis for control of the chemical feeds of lime, soda ash, and carbon dioxide. Physical titrations for caustic concentration leaving the recausticizer and calcium concentrations across the softening clarifier are periodically made. Chemical feed rates are adjusted if necessary.

IV. PERFORMANCE CHARACTERISTICS

The dual alkali system at Chevrolet-Parma was officially placed on line in early March, 1974. Since that time, it has been going

through a checkout and shakedown period to determine the actual performance characteristics of the system. During this period, many interrelated and unrelated mechanical and chemical difficulties were encountered. Some of these problems did cause complete shut-downs, but most were overcome through minor process and operational modifications while on stream. It is the purpose of this section to describe the performance of various parts of the system.

SCRUBBER PERFORMANCE

The heart of the system, of course, is the absorbers for removing the sulfur dioxide. The chemical reactions that take place in these scrubbers, however, will greatly affect the operation and economics of the regeneration section of our system. Thus it is not just a matter of scrubbing out SO_2 , but doing it with a controlled scrubber chemistry.

Sulfur dioxide tests across the scrubbers have shown removal efficiencies from 85 to 95% over the range of scrubber outlet pH of 5.5 to 6.5. Particulate loadings leaving the scrubbers are still to be determined. Testing is being continued to determine tray efficiencies for possible optimization.

As stated previously, scrubber chemistry will greatly affect the regeneration section of our system. Specifically, if the oxidation rate (Reaction 3) in the scrubber is too high, not enough bi-sulfite (Reaction 2) will be formed to take care of the calcium carbonate (Reaction 9) in our neutralizer reactor. This means that the lime utilization of the system would be very poor. For complete utilization of the lime, the maximum oxidation rate must be near 10%. Actual tests, however, show that our oxidation rates fluctuate between 30 and 80% with the recycle pH and steaming rate. The exact effect of these variables on the oxidation rate will be determined in hopes of controlling a lower rate.

In addition, oxidation rate will greatly affect the scaling tendencies in our scrubbers. Again, a sufficient amount of bi-sulfite must be present in the recycle liquor to partially neutralize the incoming scrubber feed over the top tray. If the pH on the top tray is not kept near 9.0, soluble calcium can react with the CO_2 in the flue gases and form a carbonate scale, causing high pressure drop and poor liquid-gas contact. Initially, this problem was very acute and necessitated frequent shutdowns for manually removing the calcium carbonate scale. Since then, however, revisions have been made to alleviate this problem. First, a

$$\text{*Oxidation Rate} = 100\% - \frac{\text{Total normality of HSO}_3 + \text{SO}_3 \text{ in recycle}}{\text{Normality of OH in scrubber feed}} \times 100\%$$

pipng revision was made to assure better mixing of the acidic recycle liquor with the incoming scrubber feed on the top tray. Second, the recycle liquor is now controlled at a 6.0 pH instead of 6.5 pH to assure more bi-sulfite. Finally, the soluble calcium levels in the feed stream are maintained around 300 ppm through softening. The combination of these changes has now allowed us to run over a month on a scrubber without any appreciable increase in pressure drop. It is still too early to tell whether we have eliminated the scaling altogether or are just retarding it.

REGENERATION PERFORMANCE

The main function of the regeneration section is to produce a 0.1M OH⁻ solution. No problems have been encountered in doing this, but, because of the inability to reuse the CaCO₃ in the neutralization reactor, the overall lime utilization has been very poor. The lime/sulfur stoichiometry has been in the range of 1.4 to 1.5/1.0. Original design expectation were 1.1/1.0.

The problem in using the calcium carbonate have been both mechanical and chemical. As explained previously, the high oxidation in the scrubbers rate causes a deficiency of bi-sulfite in the scrubber bleed liquor. This deficiency then requires that only a small volume of carbonate slurry be metered to the neutralizer tank. This necessitates only cracking the control valve for the low flows. The slurry then has a tendency to plug the control valve. Thus we have been forced to over-feed the carbonate slurry. Also working against the neutralization is the fact that we utilize a back mix reactor. In essence, we are trying to react an acidic stream with a basic stream in an already highly basic media. A plug-flow reactor arrangement is being investigated. Once these problems are overcome, our lime to sulfur stoichiometry should improve to be 1.3:1.

PRIMARY CLARIFIER PERFORMANCE

From the regeneration section, the recausticized solution flows to a primary clarifier for solid-liquid separation. If not operated correctly, we found that this unit could cause major problems and shutdowns.

The first problem was encountered early in the debugging period. The total system was being operated on a Monday through Friday basis with inspection scheduled for the weekends. At that time, it was standard practice to shut off all pumps and completely stop the solution flow. During one of these periods, the slurry pumps at the bottom of the clarifier broke down and we could not recirculate or draw off sludge. As a result of the complete quiescence, the sludge blanket in the clarifier, approximately

60 tons, settled rapidly to the clarifier bottom and jammed the clarifier rake. It became necessary to completely drain the clarifier, manually remove most of the sludge and manually free the rake. One week was lost because of this. Now it is a standard practice to always keep an upflow through the clarifiers to keep the solids in suspension.

The second major problem occurred when we progressed from scrubbing one boiler to three boilers. After operating about a week on three boilers, solid carry over became evident, upset the softening clarifier, and got into the scrubbers causing scaling and plugging. At that time, it was not known whether this carry over was due to a change in scrubber chemistry or an inability to remove solids on a regular basis.

Originally, we thought that additional scrubbers that we placed on line had low oxidation rates and their spent solution was forming calcium sulfite (Reaction 5) when regenerated. The $\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$ is much more difficult to settle and filter than the $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$. If allowed to build up then, the calcium sulfite could form an hindered settling zone in the clarifiers. The oxidation rates, however, in all the scrubbers were found not to vary significantly.

Also, at the same time we placed the system completely on line, we encountered some scheduling problems with the solid waste hauler of our sludge. Since trucks were not available, we could not draw off enough sludge to keep up with its formation. As a rule of thumb, five tons of wet sludge should be removed for every ton of lime added to the system. Consequently, we started to build up a solid inventory in the clarifiers (400 tons), lost the sludge blanket, forced an hindered settling zone in the clarifiers, and pushed the solids over the top. A month was lost while cleaning the solids from the total system.

After realizing this, several operational changes have been made to assure that this situation does not develop again. First, it is imperative that we be able to remove solids at a greater rate than they are formed. Once the sludge blanket in the clarifier is formed, it must be maintained at the desired level. We now have the capability to remove sludge as required through additional trucks and manpower. Second, it is important that the sludge blanket in the clarifier be periodically checked. The operators now regularly check the sample ports on the sides of the clarifier to determine the solids concentration throughout the tank. The sludge withdrawal rate is then increased or decreased to keep the solid blanket below the clarifier cone. Finally, to improve clarity of the clarifier overflow, a coagulant

is being added in the contact zone. The optimum concentration of this additive is yet to be determined. Currently, we use about a 0.3 ppm concentration.

SOFTENER PERFORMANCE

The dissolved calcium levels leaving the primary clarifier are around 700 ppm. We have found that the calcium ion reduction in the softening clarifier can be controlled to any desired level through either the addition of soda ash or carbon dioxide (Reactions 7 & 8). Currently, we are controlling the effluent from the softener at about 300 ppm without noticeable scaling in the scrubbers. Material balances across the softener indicate almost 100% chemical utilization of Na_2CO_3 or CO_2 .

Initially, we experienced some problems in settling the precipitated carbonate. We have found it necessary to maintain a sufficient concentration of precipitates (5% slurry) in the reaction zone of the clarifier for effective seeding and particle growth. Also, we have started adding a coagulant to the reactor zone to improve the solution clarity at the overflow. Again, the optimum concentration is being determined. One phenomena that we still must face is a build-up of carbonate salts on the clarifier surfaces. As of yet, it is not severe enough to force manual cleaning.

SOLIDS HANDLING PERFORMANCE

Initially, considerable difficulties were encountered in the solids handling section of our system. Minor operational and process changes have proven successful. First, we had to determine the best percent solids at the discharge of the sludge pumps for our system. This is essential for proper solids handling because: (A) too thick a slurry can cause plugging of the pumps and subsequent sludge build-up in the clarifiers, and (B) too thin a slurry can cause poor performance of the vacuum filters. A slurry concentration of around 15% solids has been found to work very well. Second, on the vacuum filters themselves, we have switched from a 100% olefin to a 100% nylon cloth. The release characteristics of the cake have improved greatly. Last, we had to adjust the speed and vacuum on the filters to achieve a consistent cake.

Since we have been handling sludge on a more continuous basis, most of the vacuum filtering problems have diminished. Except for start-up and some random occurrences, the filter can run unattended and as designed. On an average, we have been discharging a 53% solids cake with or without cake washing. The percent

soluble in the wet cake without cake washing is about 5%. The effect of cake washing on soluble salts is being determined currently. Also, the optional water volume and arrangement for the wash sprays have to be determined to measure its effects on the system's water balance. Fortunately, contrary to original projections, we should not have to add a repuddler to aid in filter cake washing.

CONTROLS PERFORMANCE

In general, the control system has been able to function properly and to provide smooth responses to changes in boiler loads. Relatively constant scrubber bleed pH and good SO₂ removal efficiencies are easily maintained. Some revisions, however, had to be made in the control theory so that a steady operation could be assured. These changes were in the areas of lime feeding, clarifier monitoring, and pH sensing.

Originally, lime feed was somewhat difficult to automatically control because the magnetic flow meter used to monitor the slurry flow is susceptible to scaling. We were never really sure what the flow of the slurry actually was and, consequently, started to notice wild fluctuations in the alkalinity at the causticizer reactor. To circumvent this situation, we are now manually controlling the rate of dry lime addition to the system and set the slaking rate to maintain a constant level in our lime slurry tanks. Presently, we maintain a constant volumetric flow rate to the causticizer, but the percent lime in the slurry varies as we change the lime addition rate. A rough addition rate is known through experience and is adjusted after manually sampling for alkalinity in the causticizer.

Also, at start up, we thought that the torque reading on the clarifier rakes would be a good indication of sludge inventory and blankets heights. Experience has shown, however, that no direct correlation can be made. Thus we have had to implement the manual monitoring of the sludge blanket previously described.

In the most critical area, pH sensing of the recycle stream, we have found that the sensors have a tendency to drift and must be recalibrated every shift. This drifting may be caused by the way we are sampling rather than the instrument itself. Current practice calls for comparing the solution pH at the sensor to the pH reading of a lab instrument and adjusting accordingly. The effects of shielding the pH probe from the abrasive action of the recycle stream is being investigated.

MISCELLANEOUS HARDWARE PERFORMANCE

To date, most equipment has performed as expected with no serious problems. The fiberglass reinforced piping on the scrubber feeds and recycles show no evidence of abrasion. Slurry pumps and feed pumps are also holding out well. Some plugging and scaling problems have been encountered with other pieces of equipment, however.

The largest problem has been with the lime slakers. They have required considerable maintenance and attention. Most of this, we feel, is due to the fact that we are slaking with process water. Thus we are carrying out the same reactions in our slaker as we are in the causticizer. Gypsum scale has been evident both in the slakers and the slurry piping. A modification has been made so that we can slake with city water when the water balance of the system allows. Currently, we operate the slakers half the time on city water and half on process water. The maintenance has decreased drastically since starting this and the practice of running them continuously.

Also, some plugging and scaling of the gravity lines in the system have recently become evident. The overflow line between the primary and softening clarifiers became partially plugged and had to be descaled with a high pressure water hose. The scaling in this section was probably caused by the fact that our calcium carbonate slurry is recycled through this line. A revision in this piping arrangement is planned. In addition, the overflow line between the two reactors has shown signs of scaling. The exact cause is not currently known, but calcium carbonate is again suspected. A piping revision is also planned for here. The reactor tanks themselves also show signs of scale build-up. A regular maintenance program may have to be set up to maintain the gravity flow lines.

V. PROCESS ECONOMICS

The main reason for selecting the double alkali system for Chevrolet-Parma was that it afforded an economic advantage over conversion to lower sulfur oil or burning low sulfur coal. After assessing our current cost, we feel that this advantage is still real. A precise cost estimate at this time is impossible because of fluctuating labor requirements, projected sludge handling changes, and rising chemical cost. Currently, our best estimate is \$12/ton of coal burned. This figure represents fixed cost, current chemical cost, utilities, labor, and maintenance and is broken down as follows:

<u>Chemical Costs</u>	<u>Cost/Ton of Coal (55,000 Tons/Yr)</u>	
	\$	
Lime	.98	
Soda Ash	.82	
Carbon Dioxide	.07	
Polymer	<u>.01</u>	
Sub Total		1.88
<u>Utilities</u>		
Water	.16	
Steam	.37	
Electricity	<u>1.32</u>	
Sub Total		1.85
<u>Solid Waste</u>	.81	
<u>Labor</u>	2.04	
<u>Maintenance</u>	.67	
<u>Fixed Costs</u>	<u>4.55</u>	
Sub Total		<u>8.07</u>
Total		11.80

This cost added to the \$21/ton of coal purchase price is still some \$8/ton lower than the premium price for low sulfur coal.

The cost figures for Chev-Parma have to be used with discretion and cannot be applied across the board for every one of GM's powerhouses. Each one is somewhat unique and thus capital costs for installation can vary greatly just because of the physical plant site. As an example, an estimate of installation costs for another GM plant was made and determined to be \$8 million, more than double that of Chevrolet-Parma. This higher fixed cost coupled with the plant's lower steam requirements would mean operating costs double that of Chev-Parma. For this particular site, other control alternatives may be more economical.

VI. SUMMARY

The double alkali system is still going through intensive evaluation to determine its complete capabilities. In general, the trend of the system seems to be one of steady improvement. The system's evolution has not been easy, however, and has required a concerted effort. What must be remembered is that we are working with a highly developmental system. Not all the problems have been solved, and it will take time to establish the process and equipment reliability and the exact operating costs. Hopefully, after a complete year of operating experience, these questions can be answered positively and we will have provided an economic alternative to meeting the sulfur dioxide emission requirements. General Motors will gladly keep all interested parties informed of our progress in meeting this challenge.

BIBLIOGRAPHY

1. R. J. Phillips, "Operating Experiences With A Commercial Dual-Alkali SO₂ Removal System", 67th Annual Meeting of the Air Pollution Control Association, June, 1974, Denver Colorado.
2. R. J. Phillips, "Sulfur Dioxide Emission Control for Industrial Power Plants", Manufacturing Development, General Motors Corporation.
3. D. Draemel, "An EPA Overview of Sodium-Based Double-Alkali Processes, Part I-A View of Process Chemistry Of Identifiable And Attractive Schemes", Control Systems Laboratory, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina.

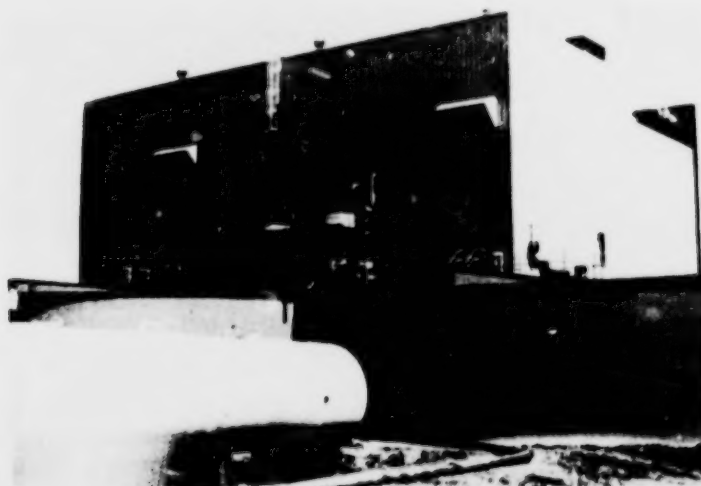


Figure 1. CHEV-PARMA
SCRUBBER BUILDING

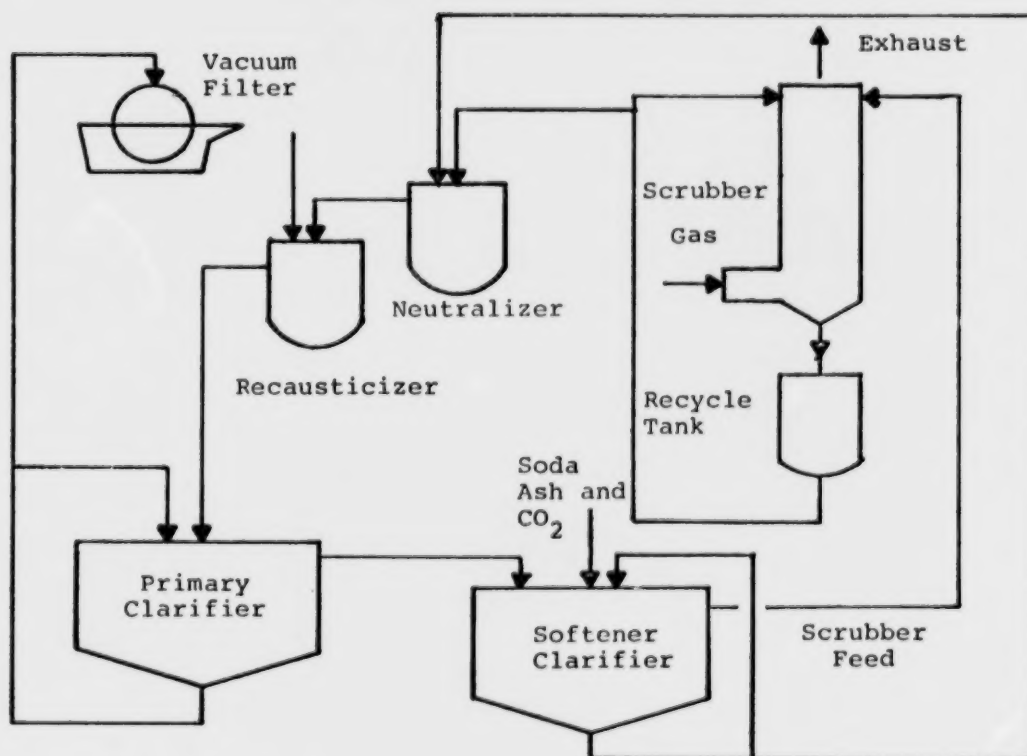


Figure 2. PROCESS FLOWSHEET

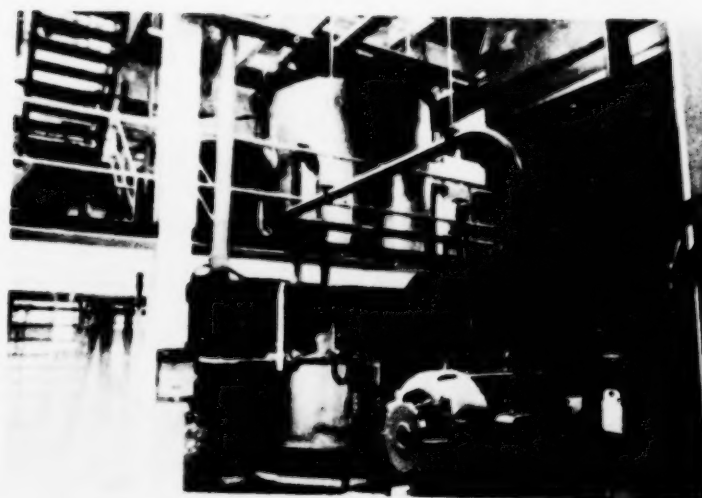


Figure 3. SCRUBBER

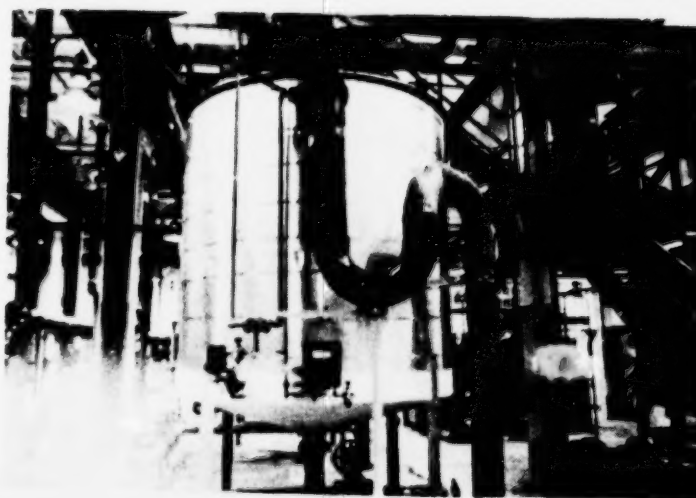


Figure 4. RECYCLE TANK

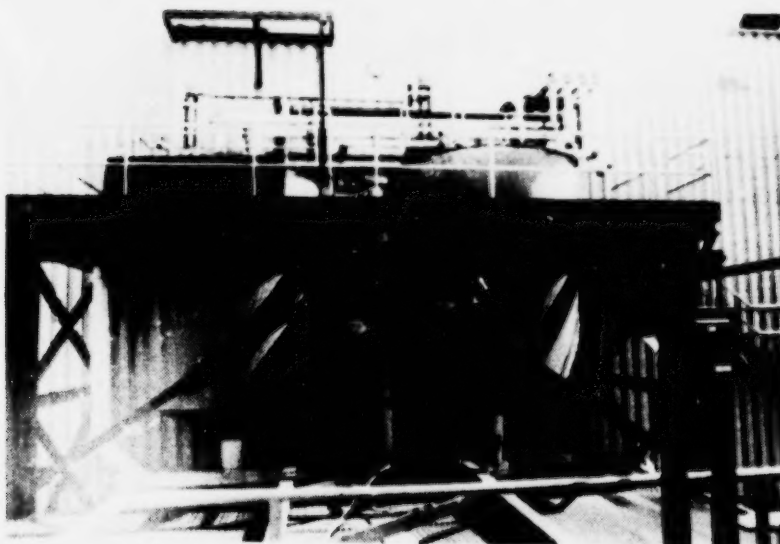


Figure 5. REACTOR TANKS

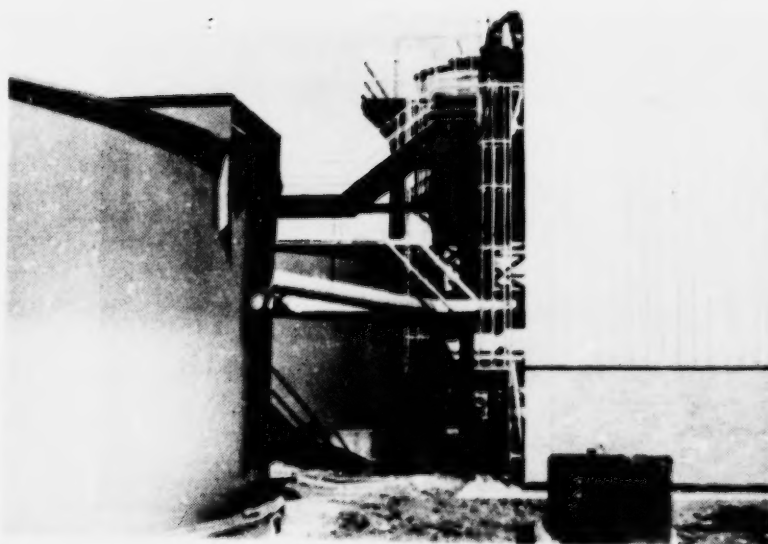


Figure 6. CLARIFIERS

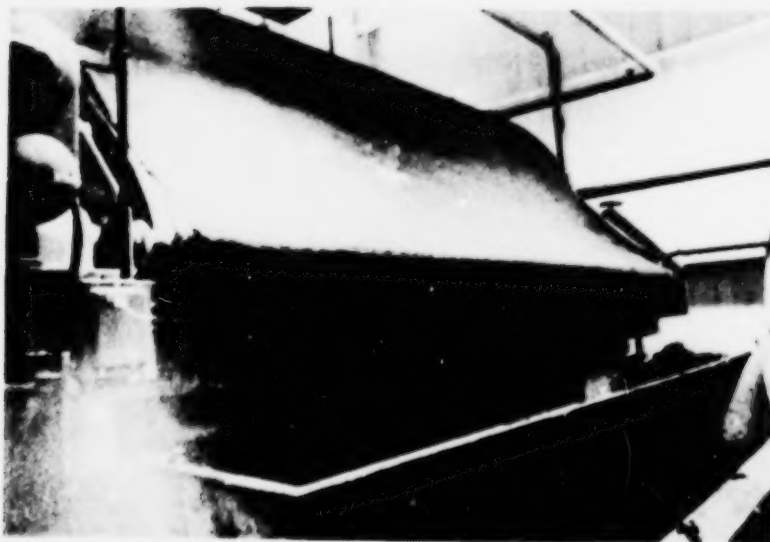


Figure 7. VACUUM FILTER

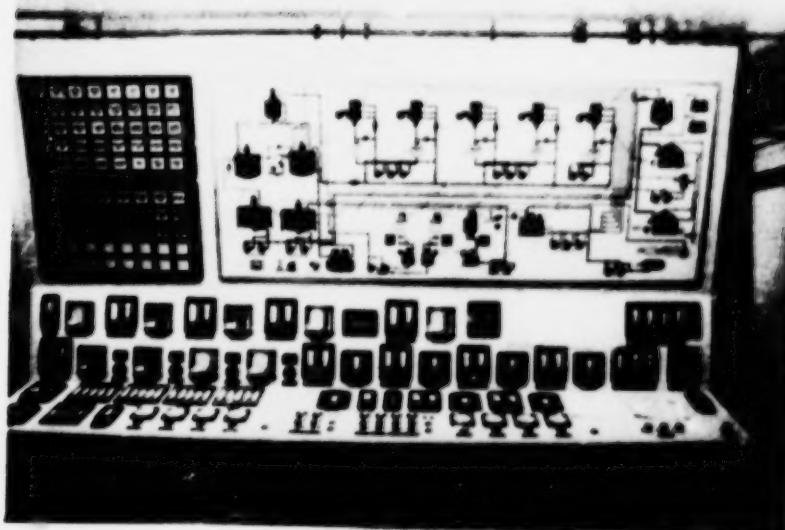


Figure 8. CONTROL PANEL

PART II

EQUIPMENT PERFORMANCE WITH
A COMMERCIAL DUAL-ALKALI SO₂ REMOVAL SYSTEM

Edmund J. Piasecki
General Motors Corporation
Argonaut Division
Detroit, Michigan

For Presentation at Environmental Protection
Agency Symposium on Flue Gas Desulfurization

Atlanta, Georgia
November 4-7, 1974

Preceding page blank

539

INTRODUCTION

Industrial boilers because of their size, load fluctuation and operating characteristics such as high excess air, relatively high dust loads, etc., make them differ from utility units.

When investigating possible control systems for SO₂ emissions, a basic criteria was established. This was as follows:

1. The system must provide complete equipment reliability.
2. The system must have a minimum of complexity.
3. The system must maintain control over a wide range of steam loads.
4. The system must be economically competitive with other control strategies.

Since this was the first full-scale installation of a double-alkali control, there was no previous experience from which to draw. However, the Argonaut Division of General Motors Corporation has had considerable experience in the design and development of systems for industrial waste treatment plants and powerhouses; and it was from this area that equipment selection and design for this plant was utilized.

EQUIPMENT CRITERIA AND PERFORMANCE

As with any new process, one can expect any number of start-up problems. These may be related to the process chemistry, mechanical equipment or a combination of both. At our Chevrolet-Parma facility we experienced some difficulties in these areas. We will point out problems along with a presentation of basic design information.

Booster Fans - The existing boiler I.D. fans were retained and the increased pressure requirements for the scrubbing system were obtained by the addition of booster fans. The fans were located upstream of the scrubbers to allow them to run in the dry inlet gas stream. This approach eliminates the need for expensive corrosion resistant materials which would be required on the outlet side of the scrubber. The design parameters for the booster fans are:

	<u>Boilers 1 & 2</u>	<u>Boilers 3 & 4</u>
Flue Gas Flow (cfm)	61,500	58,500
Static Pressure	13"	13"
Design Temperature	393°F	730°F
Gas Density	0.0474 lbs./ft. ³	0.0347 lbs./ft. ³
BHP	187	178
Motor HP	200	200
Fan & Motor Speed	1180 RPM	1180 RPM
Electrical Characteristics	460-3-60	460-3-60

Scrubbers - After evaluation of many scrubbers, it was felt that the impingement type unit would be the unit most likely to succeed in a double-alkali type process.

This type was selected because of the need for removal of both SO₂ and particulate and its resistance to plugging. Packard towers were considered but were eliminated because of a concern for plugging which was evidenced in the pilot plant operation. Venturi scrubbers were eliminated because there was doubt about their ability to absorb SO₂ gases. Subsequently, however, a pilot operation has demonstrated the ability of the venturi unit to absorb SO₂ gases.

Basic design parameters are as follows:

		<u>Scrubbers 1 & 2</u>	<u>Scrubbers 3 & 4</u>
Flue Gas Rate	Peak	61,500 cfm	58,500 cfm
	Minimum	20,000 cfm	19,000 cfm
Gas Inlet Temperature		393°F	550°F
Gas Outlet Temperature		137°F	152°F
Particulate Loading			
(lbs. part./1000 lbs. gas)		0.6 Inlet	0.6 Inlet
		0.1 Outlet	0.1 Outlet
SO ₂		1000-3500 ppm	1000-3500 ppm
Turn Down		3 to 1	3 to 1

Scrubber material:

Non-Welded Components	316 Stainless Steel
Welded Components	316 L Stainless Steel
Mist Eliminator (Fleximesh Type)	316 Stainless Steel
Shell Thickness	11 Gauge

The scrubbers, to date, have been operating at an SO₂ removal efficiency above 90 percent.

We have experienced some problems with CaCO₃ scaling on the top tray of the unit. Scaling is caused by the following two conditions:

1. High pH condition on the top tray.
2. Calcium level in the scrubber feed in excess of 250 ppm Ca++.

The top tray pH is approximately 9 and is the result of the mixing of two separate streams on this tray; scrubber recycle pH equals 6.5 and scrubber feed of pH equals 13.

CaCO₃ precipitation is pH dependent; in general, the lower the pH the more soluble CaCO₃ becomes. At a pH equal to 7.0, we feel it would be possible to operate at a Ca++ hardness of approximately 600 ppm.

We are currently preventing scrubber scaling by giving careful attention to softening of the feed stream.

To date, we have seen no corrosion problems in the scrubber units.

Chemical Mix Tanks - Both chemical mix tanks Nos. 1 and 2 are constructed of carbon steel. The function of these tanks is to mix a slurry of slaked lime with the spent scrubber liquid. Design residence time is seven minutes at peak flow, and turbine type agitators with low RPM and high pumping rate have been provided.

Over a period of months, CaSO_4 scale has formed in the ten inch line between the two mix tanks. Since this is a gravity line, we are considering installing a flume in place of the solid pipe. This arrangement should provide easier scale removal.

Reactor Clarifier No. 1 - The unit is the reactor clarifier type which means there is a center reaction well. Mixing in the wall is developed by a turbine-draft type pump. This arrangement permits recycling settled sludge which acts as seed material for further crystal growth and also allows the addition of flocculation aid which may be necessary for good settling.

The primary function of this unit in our process is to settle CaSO_4 and CaSO_3 ; this is formed in the chemical mix tanks by lime addition.

Basic design criteria is as follows:

Design Flow	1200 GPM
Diameter	60 ft.
Liquid Depth	14 ft.
Rake Drive HP	1-1/2
Surface Rise Rate	0.46 GPM/ft. ²
Rake Torque Rating	19,000 ft./lb.
Mixer HP	3
Mixer Speed	4.9 RPM Minimum 14.5 RPM Maximum

Our experience has shown that CaSO_4 forms a large crystal and settles readily. However, CaSO_3 precipitates as a finely divided crystal; for this reason, settling rate is extremely slow. In order to operate this reactor clarifier without overflowing solids, it is necessary to add a high molecular weight anionic polymer to the reaction well at a concentration of 0.25 ppm.

To aid solids removal, this unit is equipped with sludge rakes and a cone bottom sloped at 12:1. Solids removal is started when the solids concentration in the bottom of the clarifier is 16 percent. When solids concentration reaches 20 percent, the sludge no longer flows freely; at 30 percent the sludge ceases to flow. In our case, when concentration of solids reached 30 percent, the torque rating of the unit was exceeded and it was necessary to drain and clean out the unit.

Reactor Clarifier No. 2 - The primary function of this unit is to soften the regenerated caustic stream from 800 ppm Ca++ to 200 ppm Ca++. This is accomplished by adding Na₂CO₃ and CO₂. The Na₂CO₃ is added to the reaction well as a solution; CO₂ (gas) is added using a diffusion ring located eight feet below the surface of the liquid in the reaction well.

Here, as in the case of the first unit, an anionic polymer is added to the reaction well at a concentration of 0.30 ppm to aid solids settling.

The basic design criteria for this unit is the same as for reactor clarifier No. 1.

Vacuum Filters - These units are the rotary vacuum filter type.

The basic filter criteria is:

Surface Area	500 ft. ²
Drum Diameter	10 ft.
Drum Face Length	16 ft.
Belt Material	Nylon
Drum Drive Speed	0.06 RPM Minimum
	0.5 RPM Maximum
Vacuum Receiver Size	48" Diameter X 84"
Vacuum Pump Capacity	1000 cfm at 22" Hg
Filtrate Pump Capacity	300 GPM at 15 ft. Head

The system has two vacuum filters, each capable of handling our process solids load.

Originally, the filter belt supplied was made from polypropylene cloth. However, after testing other materials, better cloth release was obtained by changing to a nylon material.

To date, it appears that the nylon material is not wearing as well as the polypropylene. This wear seems to be caused by fly ash particles which are not being washed off the cloth. We are now in the process of modifying the belt washing system in hopes of increasing belt life.

CONTROL PHILOSOPHY

When looking at possible systems for control of our new process, some basic criteria for evaluation had to be established. The following items were considered basic to satisfactory system operation:

1. The controls must be reliable.
2. Instruments must be easy to maintain and be of such type that powerhouse maintenance personnel would feel comfortable with them.
3. Control system must operate with a minimum of operator manual control.

Our conclusion was that a pneumatic instrumentation system fills these basic needs.

The control system can be broken down into three basic areas:

1. Caustic feed to the scrubber.
2. Chemical feed for caustic regeneration and softening.
3. Sludge removal and filter cake handling.

Caustic Feed - Caustic feed to the scrubber is controlled by two inputs into the control system:

1. Steam flow from the boiler.
2. Scrubber pH.

By assuming a fixed percentage of sulfur in the coal and Btu content per pound of coal, we can arrive at an approximate amount of NaOH to be added. Of course this value is only approximate because the Btu and sulfur content of the coal can and does vary.

Therefore, these variations must be accounted for. This can be done by measuring scrubber pH.

If the coal has more sulfur than normal for a given steam load, then the amount of caustic being fed to the scrubber will be insufficient to neutralize the acid generated. This will cause the scrubber pH to be low. Sensing this, the controller will increase caustic to the unit by opening the control valve a little more.

A decrease in sulfur content would similarly cause a decrease in caustic feed to the scrubber.

We control scrubber pH at 6.0 to 6.5; this yields SO₂ scrubbing efficiency in the 90 percent range.

Chemical Feeds - Once the amount of NaOH is known, then the total pounds of sulfur being scrubbed is also known. Therefore, by measuring continuously total system caustic feed rate, chemical reactants can be ratioed and controlled automatically. For example, our current lime control is set at 3 pounds CaO per 100 gallons of caustic scrubber feed used. If the gallons of caustic used goes up, the lime feed automatically increases also. NaCO_3 , etc., are also controlled in this manner.

Sludge Handling - Sludge is currently removed from the bottom of both reactor clarifiers. The bottom stream from the No. 1 clarifier contains CaSO_3 , CaSO_4 and fly ash. This clarifier is equipped with sludge rakes. Sludge blowdown is also controlled by ratio to scrubber feed.

Checks on sludge density, sludge level and rake torque determine when sludge should be taken out of the clarifiers. Sludge level and density are determined by manual testing while rake torque is read from a meter.

The same method is used for removing CaCO_3 sludge from the bottom of the second reactor clarifier.

The sludge from both reactor clarifiers is then filtered to remove water and valuable salts. The rotary vacuum filters used are essentially operated manually.

SUMMARY

The dual- or double-alkali process is still a highly developmental system and will continue to be so until the questions of performance reliability and cost are answered.

We have had some process difficulties which have been solved and a few which are under intensive study but still remain a problem.

A one-year in-depth evaluation of the total system is now in progress.

It should be pointed out that this system was designed for General Motors industrial type boilers.

Utility boilers are different in their firing and operating conditions. Therefore, direct application of industrial boiler SO₂ emission control technology would not necessarily apply to utility boilers.

REFERENCES

1. R. J. Phillips, "Sulfur Dioxide Emission Control for Industrial Power Plants," Manufacturing Development, General Motors Corporation.
2. J. H. Frazier, "A System for Removal of Sulfur Oxides from Industrial Boiler Flue Gases," Industrial Coal Conference, Purdue University, October 7, 1971, Plant & Environmental Engineering, General Motors Corporation.
3. J. H. Frazier, "Status of Sulfur Dioxide Control in Boiler Flue Gas at General Motors Corporation," Industrial Fuel Conference, Purdue University, October 3, 1973.
4. R. J. Phillips, "Operating Experiences with a Commercial Dual-Alkali SO₂ Removal System," 67th Annual Meeting of the Air Pollution Control Association, June 9-13, 1974, Plant & Environmental Engineering, General Motors Corporation.

EPA-ADL DUAL ALKALI PROGRAM--INTERIM RESULTS

by

C. R. LaMantia
R. R. Lunt
J. E. Oberholtzer
E. L. Field

Arthur D. Little, Inc.
Cambridge, Massachusetts

N. Kaplan

EPA Control Systems Laboratory
Research Triangle Park, North Carolina

For Presentation at:

EPA FLUE GAS DESULFURIZATION SYMPOSIUM

ATLANTA, GEORGIA

NOVEMBER 4-7, 1974

ABSTRACT

This paper presents interim results of the Dual Alkali Program being conducted by Arthur D. Little, Inc. (ADL), for EPA's Control Systems Laboratory. The program consists of process research and development at three levels of investigation:

- Laboratory reactor studies
- Pilot plant operations in a 1200 scfm system at ADL
- Prototype test program on a 20 megawatt Combustion Equipment Associates (CEA)/ADL system at Plant Scholz, Gulf Power Company/Southern Services, Inc.

Various modes of operating dual alkali systems on high and low sulfur fuel applications are being investigated and include:

- Concentrated and dilute sodium scrubbing systems
- Lime and/or limestone regeneration
- Slipstream sulfate treatment schemes

In each mode, the objective is to characterize the dual alkali process in terms of SO_2 removal, chemical yields, oxidation, sulfate regeneration, waste solids characteristics and soluble solids losses. This paper reports some of the laboratory and pilot plant results obtained to date on lime and limestone regeneration of concentrated and dilute sodium scrubbing solutions as well as work performed on the sulfuric acid treatment scheme for soluble sulfate precipitation.

ACKNOWLEDGEMENTS

The authors wish to acknowledge the contributions of several individuals to the EPA/ADL Dual Alkali Program and to the preparation of this paper.

The earlier part of the EPA Laboratory Program was conducted under the direction of Dean Draemel, now at the University of California. The EPA laboratory work was carried on and completed by James MacQueen and Robert Opferkuch of Monsanto Research Corporation under contract to the EPA.

At ADL, major contributions were made to the pilot plant and laboratory programs by James Valentine, Bernard Jackson and Itamar Bodek. C. Lem Kusik provided insights into the thermodynamics of the high ionic strength solutions.

The cooperation of Combustion Equipment Associates in the pilot and prototype test programs and Gulf Power Company and Southern Services in the prototype program are also gratefully acknowledged.

Finally, Frank Princiotta made important contributions in his continuing involvement in the review and planning of the EPA/ADL program.

TABLE OF CONTENTS

	<u>Page</u>
List of Tables	556
List of Figures	557
I. INTRODUCTION	559
A. Program Objectives and Outline	559
B. Description of Chemistry and Definition of Terms	560
II. LABORATORY AND PILOT PLANT SYSTEMS	565
A. Laboratory Methods	565
B. Pilot Plant	569
III. LIME REGENERATION -- CONCENTRATED MODE	575
A. Prior Work on Lime Regeneration at ADL	575
B. Laboratory Studies of Lime Regeneration of Sulfate	576
C. Pilot Plant Results	586
IV. SULFURIC ACID TREATMENT -- CONCENTRATED MODE	607
A. Laboratory Results	607
B. Pilot Plant Results	614
C. Sulfuric Acid Reactor Model	624
D. Applicability of the Sulfuric Acid Treatment System	627
V. LIMESTONE REGENERATION -- CONCENTRATED MODE	629
A. Initial Laboratory Studies	629
B. Initial Pilot Plant Tests and Subsequent Studies of Factors Affecting the Physical Properties of Product Solids	642
C. Conclusions Regarding Limestone Regeneration of Concentrated Sodium Bisulfite/Sulfite Solutions	649
VI. LIMESTONE/LIME DILUTE MODES	650
A. EPA Laboratory Results	650
B. Dilute Mode Alternatives	652
VII. SOUTHERN SERVICES PROTOTYPE TEST PROGRAM	661
A. System Description	661
B. General Outline	662
Glossary	
References	

LIST OF TABLES

<u>Table No.</u>		<u>Page</u>
II-1	Scrubber System Operating Conditions	571
III-1	Sulfate Precipitation in Concentrated Mode Lime Regeneration Laboratory Continuous Reactor-- 50-Minute Residence Time	578
III-2	Effect of CSTR pH and Hold-up Time on Solids Properties -- Concentrated Lime Mode	593
III-3	Effects on Sulfate Concentration and pH on Settling Properties of Solids Produced in a CSTR with a 30-Minute Holdup	594
III-4	Comparative Waste Solids Properties -- Concentrated Lime Mode	596
III-5	Summary of Closed-Loop Runs 402 and 404 -- Concentrated Lime Mode	601
IV-1	Summary of Laboratory Continuous Reactor Sulfuric Acid Treatment Experiments	612
IV-2	Summary of Sulfuric Acid Slipstream Treatment Results	616
IV-3	Summary of Closed-Loop Runs	621
IV-4	Model Simulations of Pilot Plant Operations	625
V-1	Lime Treatment of CSTR Effluent Slurry After Regeneration with Limestone	648
VI-1	EPA 2-Liter Reactor Continuous Fredonia Limestone Runs -- Summary	651
VI-2	EPA Runs - Potential of Line Post-Treatment Step	653

LIST OF FIGURES

Figure No.		Page
II-1	Continuously Fed, Stirred Tank Reactor System	567
II-2	Schematic Flow Diagram of Dual Alkali Pilot Plant	570
II-3	SO ₂ Removal vs. Feed Stoichiometry (Venturi & 2 Trays)	573
III-1	Sulfate Precipitation at Various Na ₂ SO ₄ Concentrations	580
III-2	Sulfate Precipitation for Various Degrees of Regeneration Liquor	581
III-3	Effect of Active Level on Sulfate Precipitation	582
III-4	Relation of Observed Apparent Solubility Products & Saturation Values for Gypsum	584
III-5	Relation of Observed Apparent Solubility Products & Saturation Values for CaSO ₃	588
III-6	(CaSO ₄ /CaSO _x) Ratio in Solids as a Function of Sulfate Concentrations	590
III-7	(CaSO ₄ /CaSO _x) Ratio in Solids as a Function of TOS Concentration	591
III-8	Average Wt% Insolubles in Filter Cake vs. [SO ₄ ⁼] -- Concentrated Lime Mode	597
III-9	Reduction of Solubles in Filter Cake vs. Wash Ratio	598
III-10	Continuous Run 402 - ADL Reactor (Na ₂ SO ₄ Makeup)	603
III-11	Continuous Run 404 - CSTR Reactor (Na ₂ CO ₃ Makeup)	604
III-12	Composite Pilot Plant Operations for 13% Overall Oxidation in System	605
IV-1	Schematic Flow Diagram of Pilot Plant Operation For Sulfur Acid Treatment Mode	615
IV-2	H ₂ SO ₄ Reactor Efficiency vs. Feed Stoichiometry	617

LIST OF FIGURES (CONT.)

Figure No.		Page
IV-3	Extent of CaSO_4 Under-Saturation in H_2SO_4 Reactor Effluent — Pilot Plant and Laboratory Data	619
IV-4	Composite Pilot Plant Operation -- H_2SO_4 Treatment Mode	623
IV-5	H_2SO_4 Reactor Efficiency vs. Calcium Utilization	628
V-1	Comparison of Limestone Reactivities in Batch Relations	632
V-2	Effect of Limestone Feed Stoichiometry on Regeneration Rate with Fredonia Limestone	634
V-3	Effect of Na_2SO_4 Concentration on Regeneration Rate with Fredonia Limestone	636
V-4	Effect of Temperature on Regeneration Rate with Fredonia Limestone	637
V-5	Utilization of Two Limestones in Batch and Continuous Reactors	639
V-6	Precipitation of Sulfate by Fredonia Limestone in 50-minute CSTR Experiments	641
V-7	Settling Curves for Solids Produced in 50-minute CSTR Experiments Using Fredonia Limestone	644
V-8	Effect of Soluble Magnesium on Batch Regeneration Rates with Fredonia Limestone	645
V-9	Settling Behavior Observed for Solids Produced in a 50-minute CSTR Using Fredonia Limestone	646
VI-1	EPA Limestone Runs 21-30 Extent of CaSO_4 Under-Saturation	654
VI-2	EPA Limestone Runs 21-30 Extent of CaSO_3 Super-saturation	655
VI-3	Dilute Limestone/Lime Schematic	658
VI-4	Dilute Lime/Carbonate Softening Schematic	659

EPA-ADL DUAL ALKALI PROGRAM--INTERIM RESULTS

I. INTRODUCTION

A. PROGRAM OBJECTIVES AND OUTLINE

Dual alkali or double alkali are generic terms used to describe flue gas desulfurization (FGD) processes involving absorption of SO_2 using a soluble alkali, followed by reaction of the scrubber effluent solution with lime and/or limestone, to produce a solid calcium sulfite/sulfate salt and to regenerate the soluble alkali scrubber feed solution. Many versions of dual alkali processes are under development. Most of the advanced systems utilize sodium as the soluble alkali; but ammonia-based systems are receiving increasing development effort. In addition to the type of soluble alkali used, another important difference among the dual alkali processes is the concentration of soluble alkali in the absorption/regeneration loop. The concentration and the choice of soluble alkali are importantly related to the regeneration of soluble sulfate produced in the system and to the control of soluble solids losses from the system. Many of the various schemes for dual alkali processes were described by D. Draemel and N. Kaplan at the last EPA FGD Symposium.¹

Work in the EPA-ADL Dual Alkali Program is limited to sodium-based dual alkali systems, covering a wide range of active sodium concentrations using lime and/or limestone for regeneration. Special attention is given to regeneration of sulfate and to the control of the soluble solids content of the calcium salts which are produced for disposal.

The objectives of the EPA-ADL Dual Alkali Program are the definition of promising sodium-based dual alkali modes of operation and characterization of those modes with regard to SO_2 removal, yields on reactants, sulfite oxidation, sulfate regeneration, waste solids characteristics, soluble solids losses, and overall process reliability.

The program is divided into three task areas as follows:

- Task I
 - ADL Laboratory Studies -- Performed at ADL Laboratories, Cambridge, Massachusetts -- Regeneration of concentrated sodium scrubbing solutions using lime regeneration, limestone regeneration, sulfuric acid treatment for sulfate removal, detailed characterization of the physical and chemical properties of solids produced in these regeneration modes.

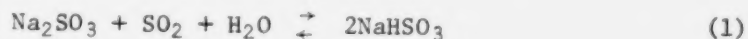
- EPA Laboratory Studies -- Performed at EPA's Control Systems Laboratory, Research Triangle Park, North Carolina -- Regeneration of dilute sodium scrubbing solutions using lime regeneration, limestone regeneration and lime/limestone regeneration.
- Task II
 - ADL Pilot Plant Studies -- Performed at CEA/ADL Pilot Facility, Cambridge, Massachusetts -- Promising modes of operation using dilute and concentrated sodium scrubbing solutions with lime regeneration, limestone regeneration, and sulfuric acid treatment for sulfate regeneration.
- Task III
 - Test Program -- Prototype Dual Alkali System at Plant Scholz, Gulf Power Company/Southern Services Inc. -- Lime regeneration of concentrated sodium scrubbing solutions to be performed on 20 megawatt CEA/ADL system.

Consistent with the program objectives, the important process characteristics will be determined for each of the dual alkali modes investigated in each task.

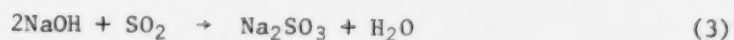
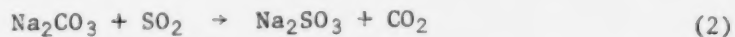
This paper reports some of the laboratory and pilot plant results obtained in the program to date under Tasks I and II.

B. DESCRIPTION OF CHEMISTRY AND DEFINITION OF TERMS

In the absorption section of sodium-based dual alkali processes, absorption of SO_2 in sodium sulfite solutions occurs to produce a bisulfite scrubber effluent solution according to the overall reaction:



Depending upon the dual alkali mode being used, the feed to the absorber may also contain some sodium hydroxide (formed in the regeneration section or used as sodium makeup) and/or some sodium carbonate (used as sodium makeup to the system). Both sodium carbonate and hydroxide form sodium sulfite on absorption of SO_2 :



which is used in further absorption to produce bisulfite. The absorber feed solution will also contain some level of sodium sulfate in solution and may contain some sodium bisulfite if neutralization is not completed in the regeneration section. Sodium, identified as associated with anions involved in SO₂ absorption reactions, is referred to as "active" sodium (includes sodium sulfite, bisulfite, hydroxide and carbonate/bicarbonate). The sulfite/bisulfite content of solutions, or total oxidizable sulfur content, is also referred to as TOS in this paper.

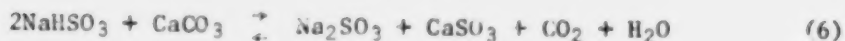
Some oxidation of sulfite to sulfate occurs in the absorber due to reaction of sulfite with oxygen in the flue gas:



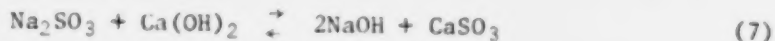
The rate of oxidation or oxygen transfer in the absorber is a function of the absorber design, oxygen concentration in the flue gas, flue gas temperature, and the nature and concentration of the species in the scrubbing solution. For a given set of process parameters, the oxidation rate in moles of sulfite oxidized per unit time is relatively independent of the SO₂ removal rate. The oxidation rate can be expressed, on an equivalent basis, as a percentage of the SO₂ removed. As an example, for a high sulfur utility boiler flue gas containing about 4-5% O₂, at about 2500 ppm SO₂ removal, approximately 10% of the SO₂ removed from the flue gas will show as sulfate in the scrubbing solution and the remaining 90% will show as sulfite/bisulfite. At steady state, this sulfate make, equivalent to 10% of the SO₂ removed, must leave the system, either as calcium sulfate or as a purge of sodium sulfate, at the rate at which it is being formed in the system.

Under similar conditions of absorber design, oxygen concentration and solution characteristics, but at a much lower SO₂ removal rate (as in a low sulfur coal application), a similar rate of oxygen absorption/reaction will result in a much higher equivalent percentage of the SO₂ removed--perhaps as high as 50% of the SO₂ removed. Also, in industrial boiler applications, where operation at higher excess air often occurs, high equivalent oxidation rates can be encountered even in high sulfur coal applications. As a consequence, in order to avoid unacceptable sodium sulfate purge requirements from dual alkali systems, modes of operation capable of precipitating both sulfite and sulfate are required; the sulfate and sulfite must leave the system as solid calcium salts with sodium sulfate and sulfite losses minimized.

The regeneration of acid sodium sulfite/sulfate scrubber effluent solutions can be considered as a sequential reaction first involving neutralization of the bisulfite using either lime or limestone:



In theory, the lime neutralization reaction should go to completion; complete neutralization of bisulfite is not possible with CaCO_3 . Using lime, the regeneration can be carried beyond neutralization to generate caustic:



to some equilibrium hydroxide concentration. The usual form of calcium sulfite produced is the hemihydrate, $\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$.

Depending upon the concentration of sulfite and sulfate and the pH of the solution, the following reaction for sulfate removal also occurs simultaneously with neutralization and regeneration reactions (5)-(7) using either lime or limestone:



The exact form of this calcium sulfate is under investigation and will be discussed later in this paper.

Thus, the level of sulfate precipitation in the overall scheme is given by the ratio of calcium sulfate to total calcium/sulfur salts produced:

$$\text{Sulfate Precipitation} = \frac{\text{mols CaSO}_4}{\text{mol CaSO}_4 + \text{CaSO}_3}$$

If the level of sulfate formation or sulfite oxidation given by:

$$\text{Sulfate Formation} = \frac{\text{mols SO}_3^{\text{=}} \text{ oxidized}}{\text{mol SO}_2 \text{ removed}}$$

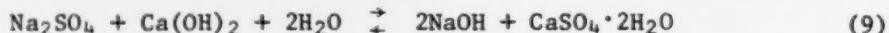
is matched by the level of sulfate precipitation, then all sulfur removed from the flue gas can leave the system as a calcium salt and no soluble sulfate purge is necessary to maintain a sulfate balance in the system. In practice, even if such a balance is established, the washed calcium sulfite/sulfate salts will contain some soluble sodium salts as well as

soluble fly ash constituents which must be purged and some sodium makeup to the system will therefore be required. Calcium utilization or yield in the overall process is defined as:

$$\text{Calcium Utilization} = \frac{\text{mols CaSO}_3 + \text{CaSO}_4 \text{ generated}}{\text{mol Ca fed}} \times 100\%$$

regardless of whether lime or limestone is used.

Dual alkali modes can be designed to accommodate varying levels of oxidation, with equivalent precipitation, up to an oxidation rate of 100%. For example, in the limit, a solution of about 0.5 molar sodium sulfate can be reacted with lime according to the following equation:



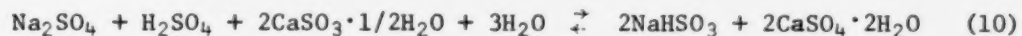
to produce gypsum and a solution containing approximately 0.1-0.15 molar sodium hydroxide and 0.45 molar sodium sulfate. If this solution, containing this relatively dilute concentration of active alkali, is recycled to the absorber where 100% oxidation recurs, then a 0.5 molar sodium sulfate solution is returned to the regeneration system and the cycle is repeated; a dual alkali system can be designed on this basis.

As the sulfite concentration in the scrubber effluent is increased, increasing the sulfite/sulfate concentration ratio in solution, the percent sulfate precipitation decreases. However, the higher concentration of active sodium species (sulfites, hydroxide and carbonates) reduces both the scrubber liquid feed rate requirement and the feed rate to the regeneration system, for a given SO_2 removal rate. As the concentration of active sodium increases, the regeneration system size is thereby reduced as are the liquid rates throughout the system. Thus, as the sulfate precipitation requirements decrease, the system size and recirculation rates can be reduced by increasing active sodium concentrations.

For purposes of this program, dilute operating modes are defined as those involving solutions containing active sodium concentrations less than or equal to 0.15 molar active Na^+ , where active sodium is considered to be sodium sulfite/bisulfite, carbonate/bicarbonate or hydroxide. Concentrated modes are those involving solutions containing active sodium concentrations greater than 0.15 molar active Na^+ .

As will be discussed in more detail, the sulfite and sulfate concentrations, besides affecting sulfate precipitation, also have a profound effect on the physical properties of the calcium/sulfur salts and, as a result, on the settling and filtration characteristics of the solids produced in dual alkali modes.

As an alternate approach to operating at decreased active sodium concentrations, high levels of sulfate precipitation can be achieved using sulfuric acid treatment for precipitation of sulfate according to the following equation:



This reaction is carried out at a low pH (2-3), where sulfite is converted to bisulfite, thereby bringing calcium sulfite into solution and exceeding the solubility product for calcium sulfate. This reaction can be applied on a process slipstream for sulfate precipitation, as required. However, in the overall process this scheme utilizes sulfuric acid, requiring additional lime or limestone for ultimate neutralization of this acidity added to the system and generating an equivalent additional amount of waste calcium/sulfur salts. This scheme is being practiced commercially in Japan and offers the advantages of operating a concentrated active sodium absorption/regeneration system with the ability to sustain high sulfate oxidation levels with equivalent precipitation of calcium sulfate.

II. LABORATORY AND PILOT PLANT SYSTEMS

A. LABORATORY METHODS

The effects of a variety of variables on the kinetics and equilibria of several different reactions related to dual alkali regeneration have been studied during the course of the laboratory experimental program. For each of the several basic reactions studied, the general approach was first to conduct a series of batch experiments in which samples of the reacting mixture were taken periodically and analyzed to determine reaction rates and final equilibrium compositions. Additional studies were then carried out in a continuously fed, stirred tank reactor (CSTR) to investigate the reaction further under conditions more representative of those in a continuous process situation.

Although a variety of reactions were studied, the experimental procedures were generally quite similar, since in most cases a solid reagent was reacted with a solution to produce a two-phase product slurry. The general experimental procedures used are presented in this section. In a few instances where a specific procedure was employed for a particular experiment, the procedural details are discussed along with the experimental findings.

1. Experimental Apparatus, Operation and Sampling Procedures

Batch experiments were carried out in a 1-liter, 3-necked, round bottom flask, immersed in a controlled-temperature water bath. A magnetically driven stirrer provided mixing. To prevent oxidation during the course of the reaction, a slow constant purge of nitrogen was maintained over the reactor headspace.

Batch experiments were initiated by weighing appropriate amounts of the water soluble reactants (reagent grade materials were used exclusively) into the flask, adding the required amount of water, initiating the nitrogen purge, and stirring to effect dissolution. When dissolution was complete, a sample of the solution was first pipetted from the flask for analysis to establish the solution composition accurately, and then the solid reactant was added to the solution to initiate the reaction. Periodically, the reactor was opened momentarily to remove a sample of the reaction slurry for analysis; procedures used for phase separation and analysis are described below.

The CSTR system used was a rather simple apparatus shown schematically in Figure II-1. The reactor was a 1.5-liter glass vessel, commonly known as a resin kettle, fitted with an overflow sidearm positioned to yield a reactor volume of about 1 liter. Mixing was performed with an electrically driven stirrer, operating at about 700 rpm. Four rectangular pieces of Plexiglas (T.M. of Rohm & Haas Co.) about 2 cm wide and 0.3 cm thick, installed vertically around the sides of the reactor, served as baffles to ensure good vertical mixing. The lower portion of the reactor was immersed in a constant temperature bath for temperature control.

Reactant feed lines, a thermometer, an inlet for nitrogen purge, and the stirrer shaft all entered the reactor through air-tight, ground glass joints in the lid; the lid and reactor bottom had mating ground glass surfaces which could be assembled air tight. During operation, the reactor headspace was purged with a flow of nitrogen which exited via the overflow sidearm.

Reactants were prepared and fed from two, stirred, 14-liter glass vessels with hermetically sealed stainless steel tops. The solid reagent was slurried with water, and fed from one tank (baffled) and the sodium sulfite/bisulfite/sulfate solution from the second. Reactants were pumped at controlled rates by variable speed peristaltic pumps (Sigma motor T6S and Masterflex 7545).

2. Analytical Procedures

The feed and effluent samples were collected and analyzed by procedures which permitted mass flow rates into and out of the reactor to be determined for each of the various chemical species of interest. Samples of solutions and slurries were collected for known periods of time in tared flasks and the total mass collected was determined gravimetrically. The slurries were then filtered, and the solids washed thoroughly with a saturated calcium sulfate solution. The weight of the resulting wet cake was determined and a representative portion taken and dried at 85°C. From the weights determined above, the measured weight loss upon drying, and the solution density, the mass flow rates of both phases of the slurry could be calculated.

Aliquots of liquids and dried solids were characterized by a variety of analytical techniques, most of them simple gravimetric or titrimetric procedures. The analytical methods employed were:

<u>Parameter</u>	<u>Method</u>
Sulfur (IV), TOS	Iodometric Titration
Total Sulfur (Solids)	Oxidize, Weigh as BaSO ₄
Total Sulfur (Solutions)	Titrate with Lead Perchlorate
Acidity/Hydroxide	NaOH/HCl Titration

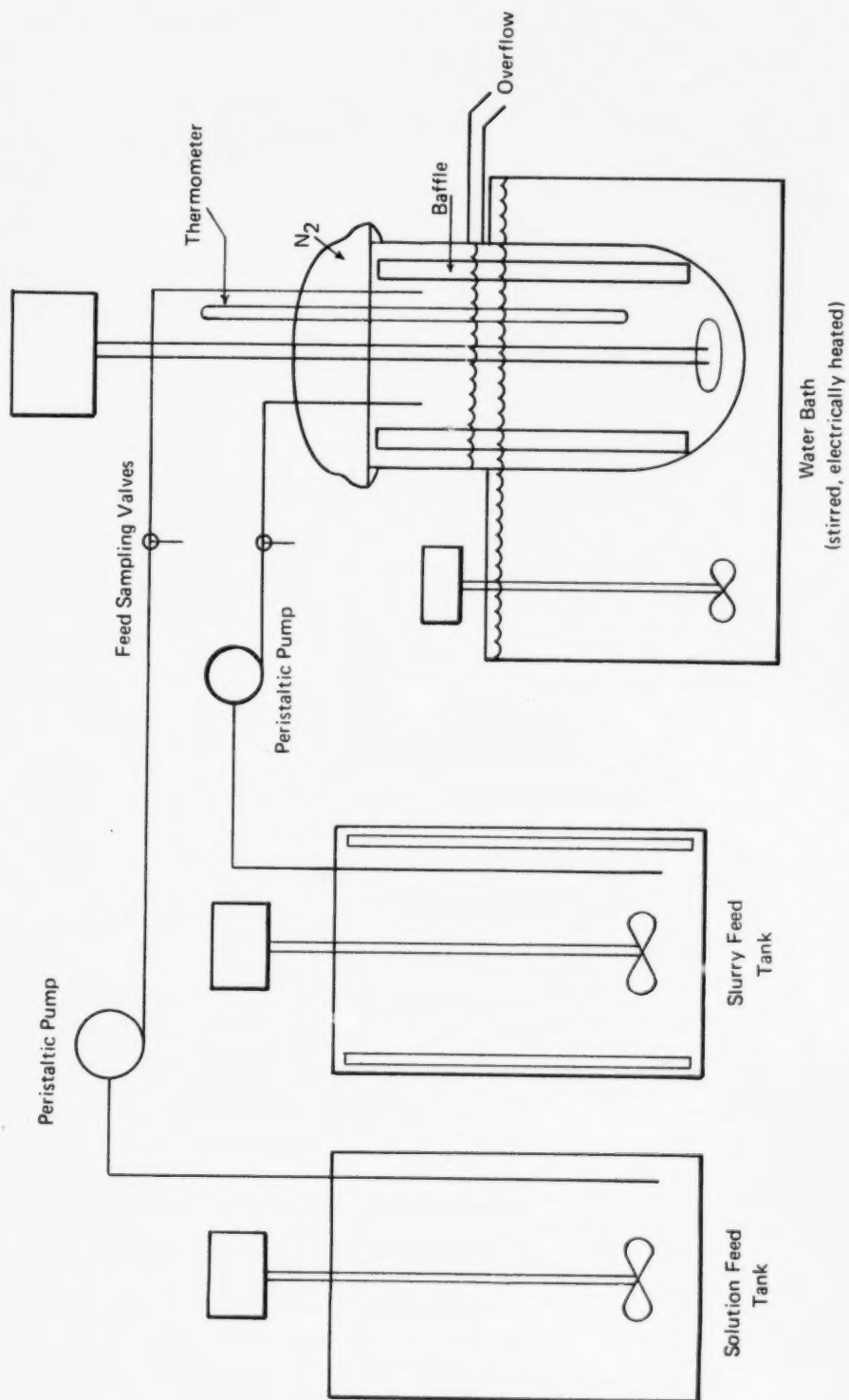


Figure II-1 Continuously Fed, Stirred Tank Reactor System

<u>Parameter</u>	<u>Method</u>
Soluble Carbonate	Titration of HCO_3^- with HCl
Solid Carbonate	Acidify, Collect CO_2 , Titrate as above
Calcium	EDTA Titration
Mg^{++} , Na^+ , (trace Ca^{++})	Atomic Absorption Spectrometry

The complete details of all analytical procedures employed are available upon request from the authors.

Determination of the amount of Na_2SO_4 which reacted with lime or limestone to precipitate CaSO_4 from sodium sulfite/bisulfite/sulfate solutions was important in many of the laboratory experiments conducted. This determination required that the filtered product solids be washed thoroughly to remove entrained mother liquor without dissolving CaSO_4 from the solids. It was for this reason that a wash solution saturated with respect to $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ was used for washing the solids. Prior to using this washing procedure routinely, analyses of dissolved calcium in the fresh and used wash solutions were performed to ascertain that no additional calcium sulfate was being introduced into the solids by the washing procedure.

The washed solids were subsequently analyzed for total sulfur, TOS, calcium, and sodium to permit the calculation of the amount of CaSO_4 present. Even after thorough washing (1 gram of dry solids washed with four, 25-ml portions of wash solution), a small amount of sodium, usually on the order of 3 mole percent of the total sulfur present, remained in the solids. That sodium was attributed to Na_2SO_4 entrapped within the calcium sulfite crystals, and it was excluded when calculating percent calcium sulfate in the solids. The percent calcium sulfate in the solids was calculated from the number of millimoles of CaSO_x and CaSO_3 per gram of total solids according to the relation

$$\text{Sulfate Precipitation} = \left[\frac{\text{CaSO}_x - \text{CaSO}_3}{\text{CaSO}_x} \right]$$

Because this calculation involves taking a relatively small difference between two larger numbers, the result is quite sensitive to small analytical errors in the determination of CaSO_x and CaSO_3 . For example, in one actual case where a calcium sulfate content of 9.4% was observed, a 1% increase in the amount of CaSO_x accompanied by a 1% decrease in the amount of CaSO_3 would have changed the calculated calcium sulfate content to 11.2%. Analytical uncertainty of that order is not unreasonable and its effect must be kept in mind when interpreting the experimental findings relating to sulfate regeneration which are discussed in succeeding sections.

B. PILOT PLANT

1. Description of the Facility

The dual alkali pilot plant at the ADL complex in Cambridge, Massachusetts, was built jointly by Combustion Equipment Associates and ADL in 1973. The system operates on 2000 acfm (400°F) of the combustion flue gas from a natural gas fired hot air furnace. SO₂ is added to the gas to provide any desired SO₂ concentration up to about 4000 ppm.

The basic facility incorporates three process sections: gas scrubbing, absorbent regeneration, and solids separation. In addition, there are auxiliary process systems and equipment for special slipstream treatment operations. A schematic flowsheet of the major process units is provided in Figure II-2.

The scrubbing system includes a variable throat venturi scrubber followed by an absorption tower fitted with a radial vane demister. The absorption tower is equipped with removable trays and sprays so that it can be operated alternatively as a spray tower, a tray tower (with from one to three trays), or a deentrainment separator.

The absorbent regeneration system contains four reactor vessels ranging in size from 17 to 700 gallons. Two lime/limestone feed systems are available--one for metering regeneration reactant as a dry solid and one for feeding reactant in slurry form.

The solids dewatering and separation system includes a 6 foot diameter Dorr Oliver thickener with three overflow levels; and a Dorr Oliver rotary drum vacuum filter with an active cloth area of 4.3 feet². The filter is equipped with a wash water spray system which can be operated in a variety of spray configurations.

In addition to this basic process equipment, the pilot plant has a number of auxiliary units including two Pfaudler kettles (one glass lined and one stainless) and a 6 inch Bird solid bowl centrifuge for operating slipstream treatment processes.

2. Scrubber Operating Characteristics

The standard scrubber configuration for the EPA test program consisted of the venturi scrubber followed by the absorption tower with two sieve trays. The tray tower was operated on a once-through basis with respect to the absorbent liquor, with the flow leaving the bottom tray being sent directly to the venturi recycle tank. There was no recycle of any spent absorbent solution around the tray tower. The standard operating conditions for the scrubber circuit are provided in Table II-1.

Because the scrubber system operation does not change appreciably for different system conditions in modes using concentrated sodium solutions, the operational characteristics of the scrubber system are reviewed here rather than for each particular system operating mode.

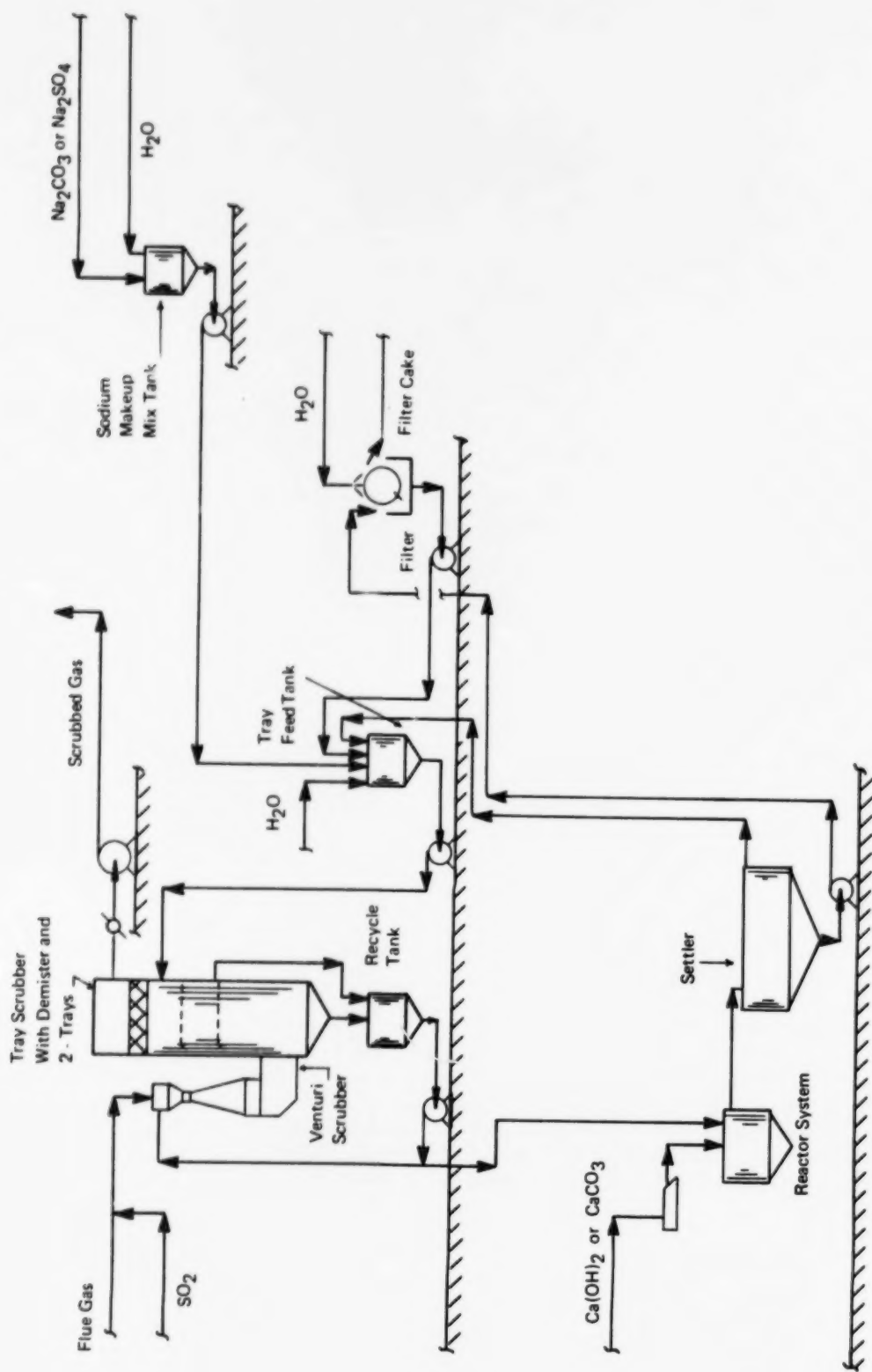


Figure 11-2 Schematic Flow Diagram of Dual Alkali Pilot Plant

Table II-1 SCRUBBER SYSTEM OPERATING CONDITIONS

Flue Gas Conditions

Temperature Entering Venturi = 390 - 430°F

Inlet Flow Rate: 1100 scfm

NO_x Level: 800 - 1000 ppm

SO₂ Level: 500 - 2700 ppm

O₂ Level: 4 - 5.5% (dry basis)

H₂O Level: 18 - 23% (dry basis)

Gas Dew Point: 126 - 131°F

Venturi Scrubber Conditions

Venturi Tank Temperature: 140 - 145°F

Venturi Recycle Flow: L/G = 16 gals/Macfm sat'd

Venturi Pressure Drop: 10 - 14 in. H₂O

Tray Tower Conditions (Two Sieve Trays)

Tray Tower Feed: L/G 1.5 - 3 gals/Macfm sat'd

Pressure Drop/Tray: 1.6 - 2.0 in. H₂O

SO₂ Removal The SO₂ removal efficiency in the pilot plant scrubber system has been evaluated in a number of previous studies as well as during this program. The overall SO₂ removal has been shown to be a function of the system feed stoichiometry (mols active Na⁺ capacity/mols SO₂ inlet), the scrubber configuration, and the inlet SO₂ level. The relationship between SO₂ removal and feed stoichiometry for the venturi two tray configuration is shown in Figure II-3. The data indicate that for a given feed stoichiometry, lower SO₂ levels tend to result in slightly lower removal efficiencies.

The plot indicates that to achieve a 90+% removal from a 2500 ppm inlet SO₂ level requires a feed stoichiometry of about 1.1; while to achieve the same 90+% removal from a 500 ppm inlet SO₂ level requires a feed stoichiometry of about 1.3.

This trend would be expected since at higher SO₂ levels there is a greater driving force for removal; hence, higher removal efficiencies are more easily attained. This is somewhat different than the experience in a direct lime/limestone scrubbing system where SO₂ absorption is limited by liquid phase and solid/liquid mass transfer. In sodium absorption above feed stoichiometries of 1.1, gas phase mass transfer appears as a limiting rate.

Scrubber Oxidation Rates Extensive tests in the pilot plant scrubber system have shown that in the concentrated active sodium ranges (0.2 - 0.5M Na⁺) the rate of sulfite oxidation (mols sulfite oxidized/unit time) is essentially constant for a given scrubber configuration and flue gas condition. This result implies that the oxidation of sulfite in the scrubber is limited by oxygen mass transfer. Thus, as the amount of gas/liquid contacting increases or as the level of oxygen in the flue gas increases, the rate of sulfite oxidation will correspondingly increase--which is in fact what happens.

Oxidation in the scrubber system plays an important role in the feed stoichiometry. Since scrubber system oxidation is virtually constant, the sulfite oxidation rates as a percentage of SO₂ absorbed (and total active sodium throughput) increase with decreasing inlet SO₂ concentration. Thus, for low SO₂ levels a greater percentage of the active sodium is converted to sulfate in the scrubber.

For the pilot plant venturi/two tray scrubber system with a flue gas oxygen concentration of 4.5 - 5.0%, the average oxidation rate is equivalent to about 240 ppm of SO₂. The stability of this oxidation rate for these conditions means that as SO₂ inlet levels decrease, the oxidation rate as a percentage of the SO₂ inlet increases.

It should be noted that these oxidation and SO₂ removal data were obtained at scrubber liquor operating temperatures which were 15 - 20°F higher than those which would be expected in a scrubber operating on a normal utility or industrial boiler. The higher temperatures were due to the fact that

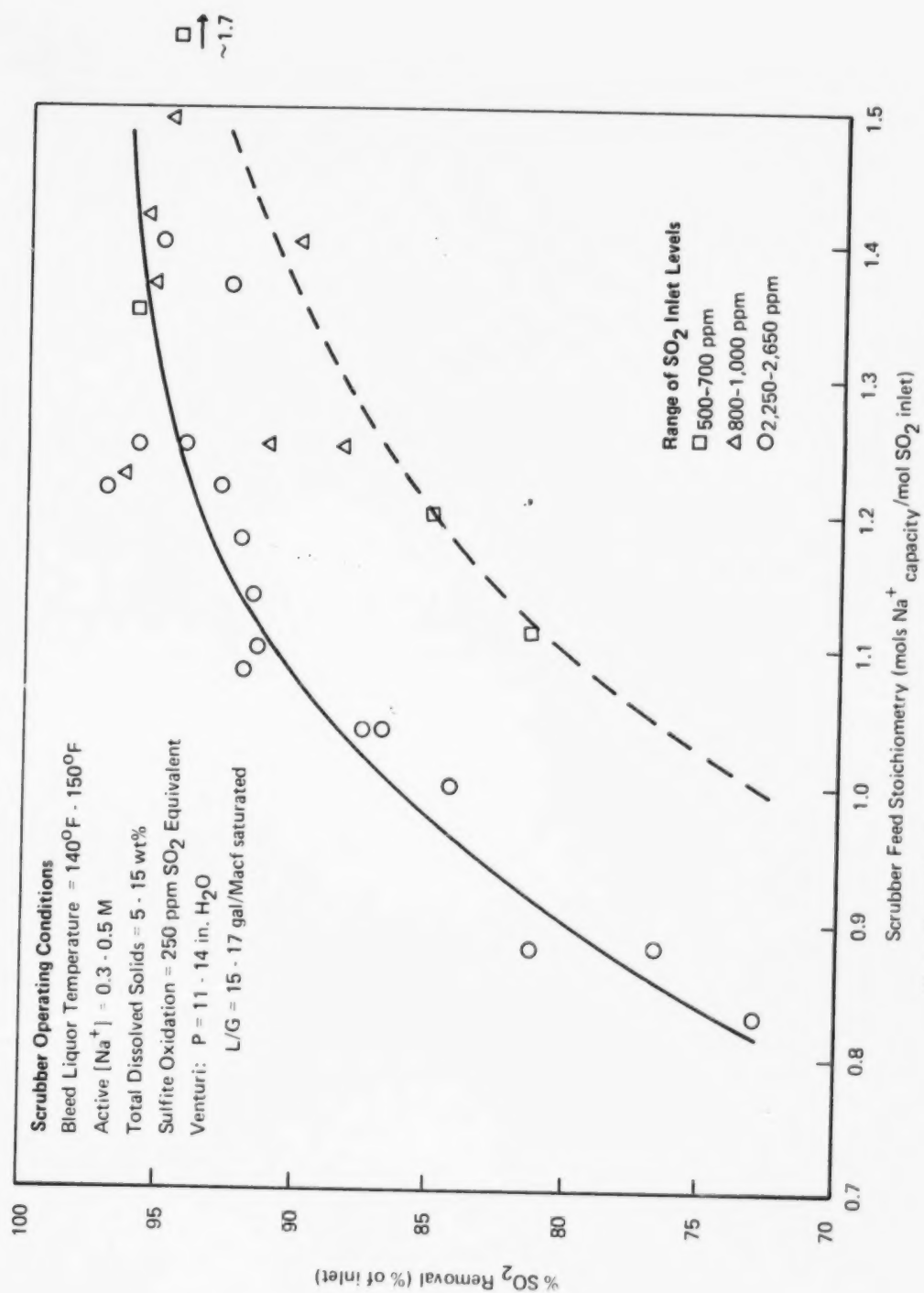


Figure II-3 SO_2 Removal Versus Feed-Stoichiometry (Venturi + 2 Trays)

the flue gas was generated from natural gas fuel which has a much lower carbon-to-hydrogen ratio than coal, and which when burned produces combustion product gas with considerably higher humidity levels. These higher humidity levels result in higher saturated gas temperatures in the scrubber which affect both SO_2 removal and sulfite oxidation rates. The equilibrium vapor pressure of SO_2 is on the order of 50 - 60% higher, so it would be expected that the removal efficiencies observed in the pilot plant would be lower than those attainable in actual boiler applications. Similarly, the rate of sulfite oxidation is strongly dependent upon temperature, increasing with increasing temperature. Thus, the oxidation rates observed in the pilot plant may also be slightly high compared to those expected in actual applications.

It should also be noted, however, that materials present in fly ash have been cited as being catalytic to oxidation in the system, and that the pilot unit used in these studies was free of fly ash. Conversely, experience has been cited in which lime/limestone systems containing fly ash have had lower oxidation rates than the same systems without fly ash. Apparently, fly ash is capable of promoting or inhibiting oxidation depending on its specific composition. However, the pilot plant was operated to produce oxidation rates on the order of those which would be expected under normal operating conditions in a full scale utility application.

III. LIME REGENERATION--CONCENTRATED MODE

A. PRIOR WORK ON LIME REGENERATION AT ADL

Prior to initiation of the EPA/ADL Dual Alkali Program, laboratory and pilot plant research and development were conducted at ADL on lime regeneration of concentrated scrubbing solutions. In 1971 ADL conducted a detailed laboratory study of the regeneration of simulated sodium scrubbing solutions using hydrated lime. This work, sponsored by the Illinois Institute for Environmental Quality,² dealt exclusively with characterizing the nature of the regeneration reaction. The objectives of the laboratory program were the development of reaction conditions which would produce waste calcium salts with good settling and filtration characteristics; production of scrubber feed solutions with low concentrations of dissolved and suspended calcium salts; and good utilization of lime in the regeneration reaction.

The objectives of these laboratory studies were satisfied and conditions for an effective regeneration reaction were determined. Chemical analyses in the laboratory were limited to liquid phase analyses; because of the generally high sulfate concentrations in the reactor feed, changes in sulfate concentration were usually within the experimental error of the analyses. As a consequence, no significant sulfate precipitation was detected in the lime regeneration of concentrated sulfite solutions.

The dual alkali process still appeared attractive because of the highly effective SO₂ removal, very low calcium concentrations in the regenerated scrubber feed, and the effectively complete utilization of lime in the process. Based upon the laboratory results, the dual alkali pilot plant was constructed at ADL's facilities. The pilot plant was operated to generate the design data for the 20 megawatt prototype dual alkali system built by CEA for Gulf Power Company/Southern Services.

As part of those pilot operations, analytical techniques were developed for the solids generated in the dual alkali system. Analyses indicated that pilot plant solids contained 10-20% of the calcium sulfur salts as calcium sulfate and that simultaneous calcium sulfate precipitation was occurring, to a limited extent, with precipitation of calcium sulfite in concentrated scrubbing solutions. Care was exercised in analysis to ensure that the sulfate found in the solids was not either occluded Na₂SO₄ or precipitated CaSO₃ which had later oxidized to CaSO₄.

The results of the laboratory and pilot work conducted prior to the EPA/ADL program are summarized in a recent paper.³

B. LABORATORY STUDIES OF LIME REGENERATION OF SULFATE

1. Introduction

The observation during ADL pilot plant tests that calcium sulfate was present in the product solids when regenerating with lime in a concentrated active sodium mode aroused a great deal of interest because in a concentrated mode the removal of sulfate produced as a result of sulfite oxidation was foreseen as a possibly significant problem. From a consideration of the thermodynamic equilibria involved, one would not expect calcium sulfate to precipitate during regeneration of concentrated sodium sulfite/bisulfite/sulfate solutions.

Consider a typical solution in a concentrated mode, regenerated to a pH in the range 9 - 11, containing 0.5M Na_2SO_4 and 0.25M Na_2SO_3 . Equilibrium levels of total dissolved calcium are typically less than 100 ppm (0.0025M). For calcium sulfate to precipitate as gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$), the apparent solubility product, " K_{sp} " (the product of total dissolved calcium and total sulfate concentrations), is about 1.05×10^{-2} . For the 0.5M Na_2SO_4 level taken as an example, a calcium level of 0.021M, about 10 times as much as is actually there, would be required to effect the precipitation of gypsum.

Subsequent to observing the presence of CaSO_4 in the solids produced at the ADL dual alkali pilot plant, Borgwardt⁴ demonstrated quite conclusively with his pilot limestone slurry scrubber that he could produce calcium sulfite product solids containing significant amounts of CaSO_4 from liquors unsaturated with respect to gypsum. He could not identify a distinct gypsum or anhydrite (anhydrous CaSO_4) crystalline phase in the solids and proposed that the CaSO_4 was present as some sort of solid solution within the $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$ crystal. He further found that he could correlate the ratio of $\text{CaSO}_4/\text{CaSO}_3$ in the solids quite well with the activities of the sulfate ion in the liquor.

The laboratory program conducted to study this apparently similar phenomenon in the concentrated lime dual alkali system consisted of a series of continuously fed, stirred tank reactor (CSTR) experiments. They were designed to first verify under "normal" operating conditions that sulfate was in fact being precipitated and then to study the effect of Na_2SO_4 concentration on the amount of sulfate precipitation. Studies were also conducted to determine the effect on sulfate precipitation of less than complete neutralization of bisulfite, regeneration with shorter and longer than normal reaction times, and higher and lower than normal levels of active sodium.

Normal reactor residence time for these experiments was established at 50 minutes, a time, based on previous ADL experience, adequate for completion of reactions, even if regeneration is carried well into the hydroxide range. The solution to be regenerated normally contained about

0.5M active sodium at a pH ranging from 5.4 to 5.6 ($\text{HSO}_3^-/\text{SO}_3^{=}$ ratio approximately 10). As noted earlier, in the laboratory apparatus lime was fed to the reactor as a 5 - 10% slurry which diluted the sodium feed by 10 - 20% upon mixing in the reactor. In the subsequent discussions, when reactant concentrations are reported, they are after dilution by the slurry water.

2. Experimental Results

The precipitation of CaSO_4 observed under normal conditions at several levels of dissolved sulfate is tabulated, along with the associated solution compositions, in Table III-1. The first two experiments, A and B, are essentially identical except for the change in concentration of Na_2SO_4 from about 0.2M to about 0.55M. In both cases measurable amounts of sulfate precipitation were observed -- precipitation increased with increasing sulfate concentration. In experiment B, regeneration was carried out to the point where the HSO_3^- fed to the reactor was just 100% neutralized and no free hydroxide was generated (sulfate precipitation was 9.1%).

Since in certain cases it proved difficult to adjust reagent flows to achieve exactly 100% neutralization of the HSO_3^- present, and in other experiments, the degree of regeneration was intentionally varied, the term "percent neutralization" has been employed to define the extent to which regeneration was carried out. When not carried to complete neutralization, the percent neutralization was computed from the relation

$$\text{Percent Neutralization} = \left[\frac{[\text{HSO}_3^-]_{\text{initial}} - [\text{HSO}_3^-]_{\text{final}}}{[\text{HSO}_3^-]_{\text{initial}}} \right] \times 100\%$$

When regeneration was carried on into the free hydroxide region, the neutralization was greater than 100% and was calculated from the relation

$$\text{Percent Neutralization} = \left[\frac{[\text{HSO}_3^-]_{\text{initial}} + [\text{OH}^-]_{\text{final}}}{[\text{HSO}_3^-]_{\text{initial}}} \right] \times 100\%$$

Experiment C was similar to B except that percent neutralization was increased from 100% to 124%. However, the sulfate precipitation (9.0%) was essentially identical to that observed at 100% neutralization. In experiment D, the concentration of sulfate was further increased to about 1.1M, and sulfate precipitation increased to 16.5%. Experiment D was carried to only 67% neutralization; however, results which will be presented later establish that Na_2SO_4 level, rather than incomplete neutralization, was the more significant factor in the increased amount of calcium sulfate precipitation observed.

Table 111-1 SULFATE PRECIPITATION IN CONCENTRATED MODE LINE REGENERATION LABORATORY CONTINUOUS REACTOR--50-MIN RESIDENCE TIME

Spt.	Composition of Feed Liquid				Composition of Reactor Effluent						° HSO ₄ ⁻ Neutralized	% Calcium Utilized ^a	moles CaSO ₄ /mol CaSO ₄ in solids
	[HSO ₄ ⁻] _{in}	[Ca ⁺⁺] _{in}	[SO ₄ ⁼] _{in}	pH	[HSO ₄ ⁻] _{out}	[OH ⁻] _{out}	[SO ₄ ⁼] _{out}	[Ca ⁺⁺] _{out}	pH				
A	0.145	0.046	0.204	5.5	0.020	nil	0.208	0.290	2.0	7.8	94	95	0.023
B	0.342	0.041	0.543	5.4	nil	nil	0.211	0.340	6.0	8.5	100	103	0.091
C	0.342	0.030	0.531	5.3	nil	0.077	0.161	0.518	5.6	11.5	124	90	0.090
D	0.154	0.030	1.110	5.4	0.116	nil	0.168	1.051	8.4	6.8	67	97	0.165

^aBased on $(\text{CaSO}_4 / \text{Ca}_{\text{total}}) \times 100$ in solids.

The sulfate precipitated in experiments A - D is shown as a function of sulfate concentration in Figure III-1. Within experimental error, sulfate precipitation was directly proportional to sulfate concentration in the reactor. (It should be remembered from the discussion of experimental error in Section II-A that sulfate precipitation is subject to some experimental error -- for a true precipitation of 8%, experimental error could cause the calculated value to be in the range from 6 - 10%.) Included in Figure III-1 are the results of additional experiments carried out at much shorter than normal residence times (5 and 8 minutes), but at the same active sodium level as experiments A - D and all at 100% neutralization. With the possible exception of the experiment at 0.2M Na_2SO_4 and 5-minute residence time, calcium sulfate precipitation and its dependence on sodium sulfate concentration was essentially the same as that observed at the normal, 50-minute residence time. Also included in Figure III-1 is the effect of allowing portions of reactor effluent slurry to stand with gentle stirring for a period of 3 hours after emerging from the reactor. In every case, a decrease in the mole ratio $\text{CaSO}_4/\text{CaSO}_x$ in the solids was observed upon standing.

The effect of percent neutralization was studied in detail with the CSTR operating at a 5-minute residence time followed by a 3-hour hold time to observe the effect of time on reaction products, equilibria, and product stability. Observed sulfate precipitation is shown in Figure III-2 as a function of percent bisulfite neutralized. In these studies, active sodium (after dilution) and the pH of the feed were maintained at "normal" levels; except for the several cases where specifically indicated, the concentration of Na_2SO_4 was between 0.64 and 0.69M. In those cases where the Na_2SO_4 concentration fell outside that range, the arrows indicate what the probable sulfate precipitation would have been at a sodium sulfate concentration of 0.66M based on the Na_2SO_4 dependence shown in Figure III-1. Although there is some scatter in the data, increased sulfate precipitation seemed to occur at very high and very low percent neutralizations. Also shown in Figure III-2 is the general decrease in $\text{CaSO}_4/\text{CaSO}_x$ in the solids when the product slurry was allowed to react for an additional period of 3 hours.

Sulfate precipitation as a function of the concentration of TOS was studied at an 8-minute CSTR residence time and the results observed are shown in Figure III-3. Again, the pH of the feed solution was maintained constant at 5.5 in these experiments. Over the relatively small range studied to date, no significant variation in sulfate precipitation was observed.

3. Discussion

The experimental work described has been aimed primarily at developing a practical understanding of what variables are important to, and in what manner they influence, sulfate precipitation. Experimental results clearly establish that calcium sulfate can be precipitated when reacting concentrated sodium sulfite/bisulfite/sulfate solutions with lime. A limited number of samples of the solids produced under a

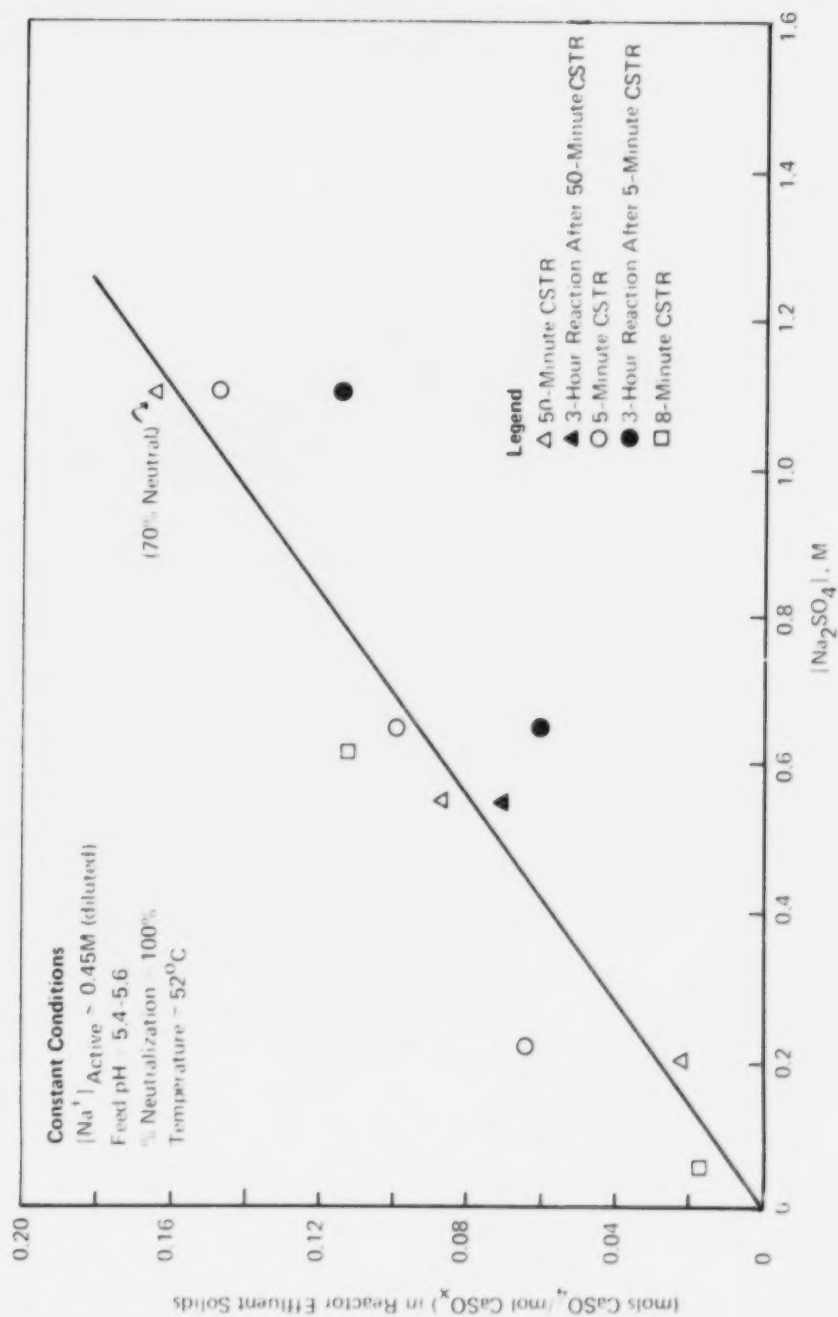


Figure III-1 Sulfate Precipitation at Various Na_2SO_4 Concentrations

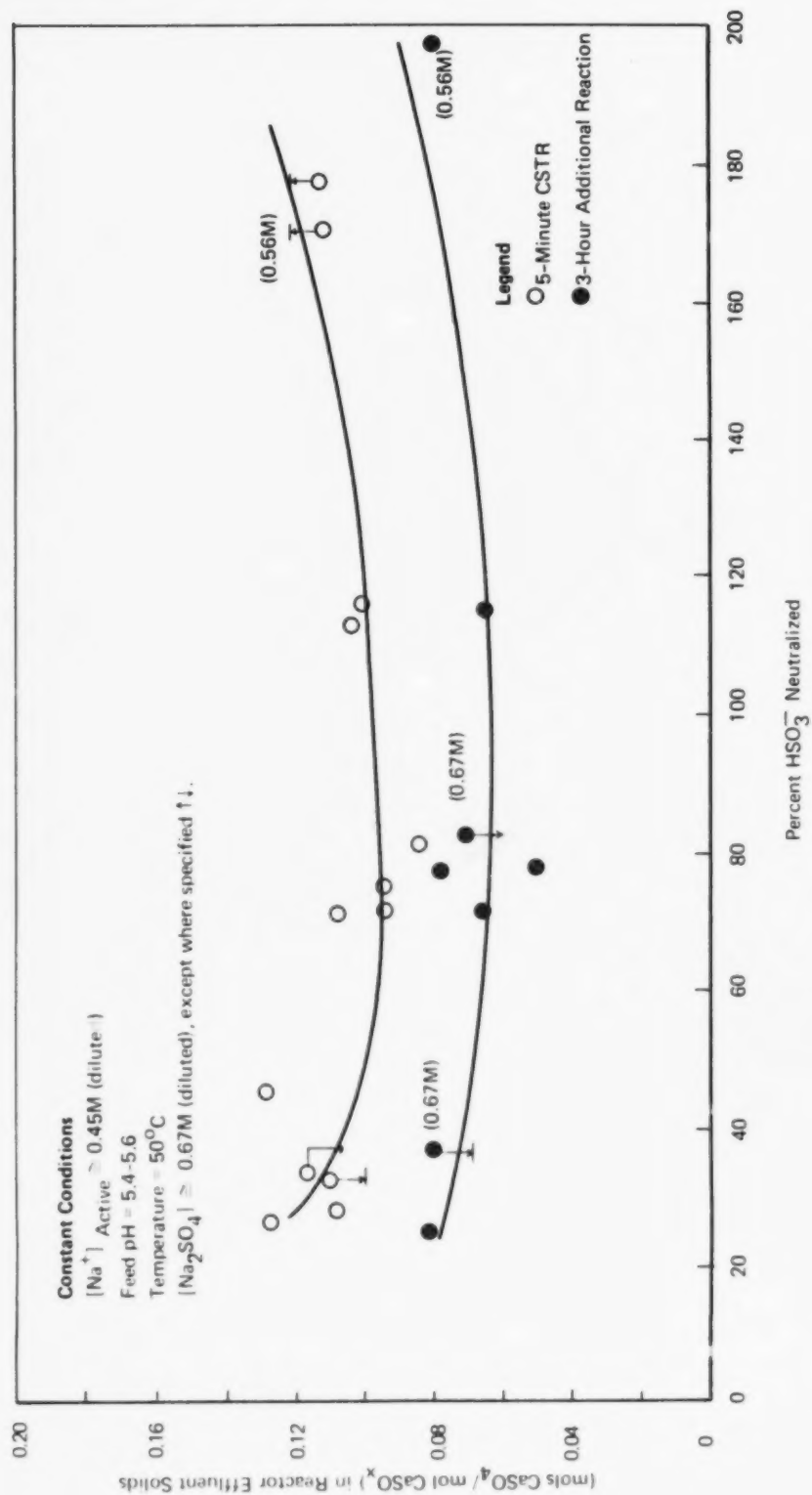


Figure III-2 Sulfate Precipitation For Various Degrees of Liquor Regeneration

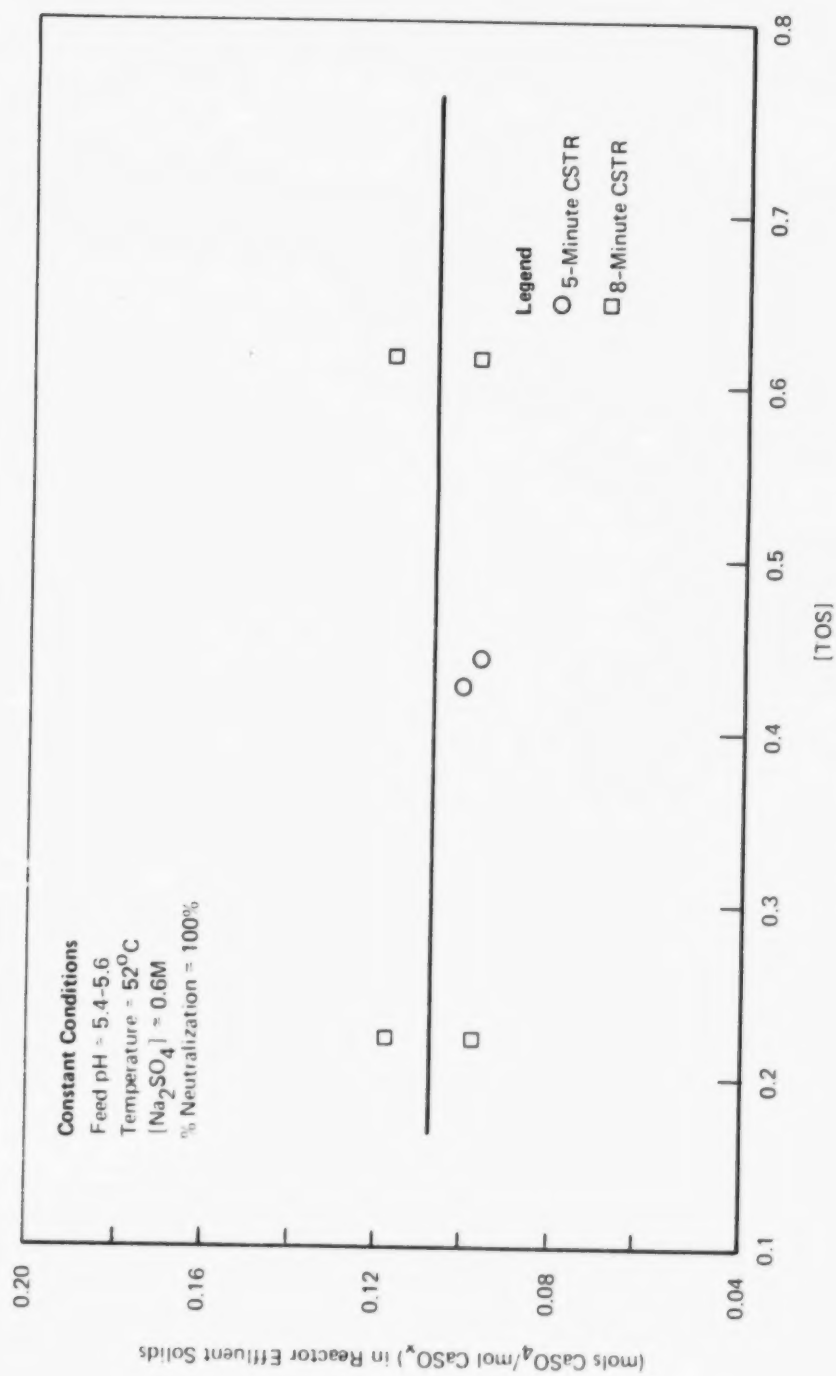


Figure III-3 Effect of Active Sodium Level on Sulfate Precipitation

variety of conditions have been examined by X-ray diffraction and the only sulfur-containing phase observed was $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$. Borgwardt found the same to be true with the CaSO_4 -containing solids produced by his direct slurry scrubbing system when operated unsaturated in gypsum; his hypothesis that CaSO_4 is in some way incorporated as a solid solution, or otherwise, in the $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$ crystal lattice seems to be applicable to the behavior observed in these studies as well.

While no attempt has been made to use rigorous solution thermodynamics to explain the effect of the several independent variables studied on observed calcium sulfate precipitation, the direct proportionality between the most significant independent variable, Na_2SO_4 concentration, and the percent CaSO_4 in the product solids is not surprising since, simplistically, increasing the $\text{SO}_4^{2-}/\text{SO}_3^{2-}$ ratio should favor precipitation (or inclusion) of CaSO_4 in the presence of calcium sulfite. The lower $\text{CaSO}_4/\text{CaSO}_x$ ratio observed at intermediate percent neutralizations is also not unreasonable because SO_3^{2-} concentration is relatively higher in that region than at either extreme. However, the observation that sulfate precipitation was not changed significantly when the concentration of TOS was changed was unexpected; at constant pH, SO_3^{2-} concentration is directly proportional to the concentration of TOS. Additional CSTR experiments covering a wider range of TOS concentrations are currently underway to define this effect better.

A general decrease in $\text{CaSO}_4/\text{CaSO}_x$ mole ratio when the effluent slurry from the 5-minute CSTR was allowed to continue to react obviously indicates that the initial effluent was not at equilibrium. The initial effluent liquid phase tended to be supersaturated with respect to CaSO_3 as evidenced by the subsequent appearance of solids (largely $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$) upon allowing samples of filtered reactor effluent to stand. However, the change in total dissolved calcium which occurred when the intact slurry samples were allowed to stand was, in most cases, too small to account for the change in $\text{CaSO}_4/\text{CaSO}_x$ observed if only a subsequent precipitation of additional calcium sulfite had occurred. Thus, one must conclude that some of the CaSO_4 initially precipitated, in fact, redissolved as the mixture was allowed to equilibrate.

In no case were the measured levels of total dissolved calcium and sulfate high enough to even approach saturation with respect to gypsum. Values of " K_{sp} ", the "apparent solubility product", which were observed under a number of reaction conditions are shown in Figure III-4. Also shown is the calculated value of " K_{sp} " as a function of ionic strength for solutions saturated with respect to gypsum. These values of " K_{sp} " were calculated by the method of Meissner and Kusik,⁵ which was developed to deal with solutions at high ionic strengths. The accuracy of the predicted values of " K_{sp} " is confirmed by the close agreement between the calculated values as shown by the curve and experimental values derived from EPA experiments in which dual alkali reaction effluent slurries were saturated with gypsum (Section VI-A) during saturation tests.

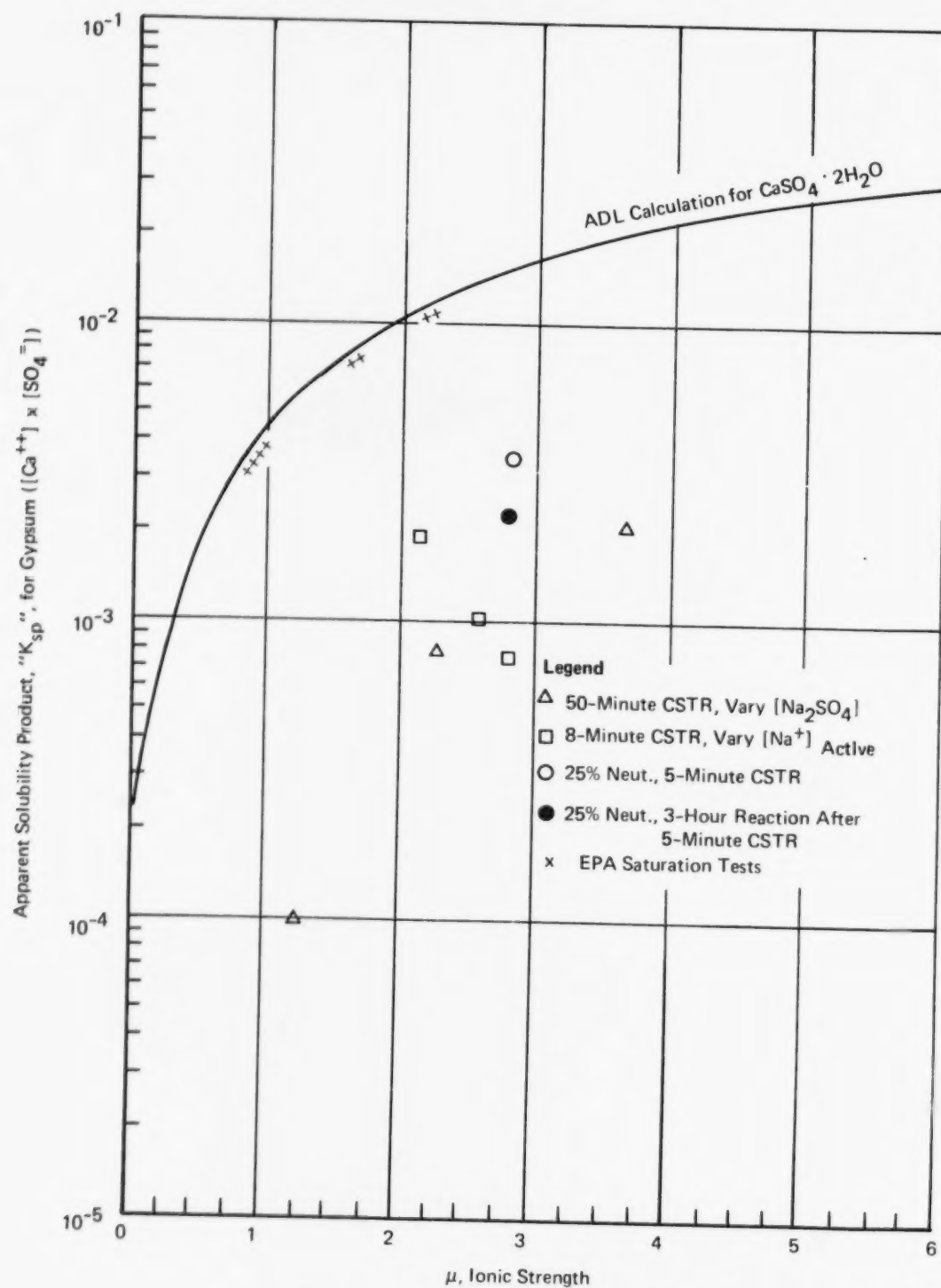


Figure III-4 Relation of Observed Apparent Solubility Products to Saturation Values For Gypsum

For a given experiment, one can express the degree of saturation of the solution with respect to gypsum by expressing the observed concentration of total soluble calcium as a percent of the corresponding calcium level at saturation taken from the curve. The percent saturation calculated for the three 50-minute CSTR experiments shown in Figure III-4 ranges from 1 to 11%, increasing with increased Na_2SO_4 concentration and $\text{CaSO}_4/\text{CaSO}_x$ ratio. The largest " K_{sp} " observed -- after 25% neutralization in a 5-minute CSTR -- corresponded to a solution only 22% saturated with respect to gypsum. Experimental studies to better understand the mechanism of CaSO_4 inclusion and its kinetic and/or thermodynamic basis are continuing in the ADL laboratory program.

C. PILOT PLANT OPERATIONS

1. Pilot Plant Test Program

The evaluation of the dual alkali process operation in the concentrated active sodium mode using lime for absorbent regeneration was focused primarily on the performance of the regeneration system. The operation of the regeneration reactor ultimately determines the required calcium-SO₂ stoichiometry as well as the properties of the waste solids produced, both of which are very important factors in determining the operability and economics of this dual alkali approach.

The pilot plant test program involved the characterization of two different reactor systems: a continuous stirred tank reactor (CSTR) and a reactor system previously developed by ADL (ADL reactor). Standard operating conditions for the ADL reactor were established in pilot plant development work prior to this EPA program; however, data on the operation of a CSTR in this mode were not available. A preliminary study of the CSTR, therefore, was undertaken in the pilot plant to examine the effects of various reactor parameters and selected process variables including sulfate and total oxidizable sulfur (TOS) concentrations in the reactor feed liquor, on the performance of the CSTR; and to develop a set of suitable operating criteria. This preliminary study consists of short-term, open-loop reactor runs using simulated scrubber bleed and dry hydrated lime. This initial phase of testing was then followed by a number of extended, closed-loop continuous runs to confirm the open-loop results and to evaluate the overall process operation using both the CSTR and the ADL reactor.

The lime used in all the pilot plant runs was a 200 mesh hydrated lime containing 95 wt% calcium (reported as Ca(OH)₂) and averaging 91 wt% available calcium hydroxide. The remaining 4 wt% calcium, consisting mainly of CaCO₃ and burnt CaO, was determined to be essentially unavailable for reaction above a pH of 8.

2. Regeneration Reactor Performance

Reactor performance in both the open-loop reactor tests and the closed-loop continuous runs was evaluated in terms of four process criteria: calcium utilization, effluent soluble calcium concentration, reaction of sodium sulfate to precipitate calcium sulfate, and the dewatering properties of the solids generated. Of particular concern in the pilot plant work were the effects of sulfate concentration in the feed liquor and the extent of absorbent regeneration (reactor pH) on the reactor performance.

The spent scrubber solution fed to the regeneration system, whether simulated or actual, normally contained 0.4 - 0.55M active sodium (0.35 - 0.50M TOS) at a pH of 5.8 - 6.1. In some tests, though, active sodium concentrations were reduced to as low as 0.25M Na⁺ (0.2M TOS). Sulfate concentrations in the spent scrubber liquor were varied from 0.05M to 1.0M SO₄⁼.

As previously indicated, all of the open-loop reactor testing as well as much of the closed-loop continuous operation was devoted to the development and characterization of the CSTR. However, in many respects the performance of the ADL reactor and the CSTR were found to be equivalent.

a. Calcium Utilization Calcium utilizations in runs with both reactor systems were high -- generally greater than 95% of the available Ca(OH)_2 in the lime feed. There appeared to be no effect of reactor configuration on the utilization of calcium; nor was there any apparent effect of sulfate concentration on calcium utilization over a range up to 1.0M SO_4^{2-} . There was, however, a slight dependence of reactor operating pH on utilization.

In the pH range 7.4 - 11.0, calcium utilizations generally approached, and sometimes exceeded, 100% of the available Ca(OH)_2 for reactor holdup times from 25 to 60 minutes. Utilizations in excess of 100% can be attributed to a number of factors including variations in the lime feed composition and analytical errors; as well as the reaction of a portion of the calcium value normally considered to be unavailable at high pH levels. The average utilization observed in all runs below a pH of 11.5 was $99\% \pm 4\%$ of the available Ca(OH)_2 , which is equivalent to approximately 95% of the total calcium in the lime, as determined from the analysis of the reactor effluent solids.

At pH levels of 11.5 - 12.5, the average utilization dropped off slightly to $96 \pm 1\%$ of the available Ca(OH)_2 (or 91-92% of the total calcium value). During continuous closed-loop operation at a pH of 11 - 11.5 and with a reactor holdup of 25 - 30 minutes, the thickener overflow and filtrate were observed to have a slightly higher pH level than the reactor effluent. This increase in pH, on the order of 0.2 pH units, indicated that the reactor effluent at the shorter residence times was not quite at equilibrium, and that some additional reaction was probably occurring in the thickener.

b. Soluble Effluent Calcium Levels The level of soluble calcium in the reactor effluent slurries for both reactor systems generally ranged from 30 - 90 ppm Ca^{++} . In only one instance did the calcium level rise above 100 ppm, to 120 ppm. Although these calcium levels were quite low, there was apparently some supersaturation with respect to $\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$ at holdup times of both 30 and 60 minutes. Figure III-5 shows the product of the calcium and sulfite concentrations measured in the effluent solution compared with the apparent solubility product curve predicted by the method of Kusik and Meissner.

The general location of these experimental solubility product data with respect to the predicted curve is consistent with data generated in both the ADL and EPA laboratory work (see Section VI. A.). Figure III-5 suggests supersaturation, however, this was not confirmed by the long-term closed-loop data. In the closed-loop continuous runs the soluble calcium levels in the thickener overflow tended to be slightly lower than or roughly equivalent to the soluble calcium levels in the reactor effluent. Since the thickener generally had a holdup time of about eight hours, it would be expected that there would be a noticeable decrease in soluble calcium levels in the overflow if the reactor effluent slurry were supersaturated to the degree indicated in Figure III-5.

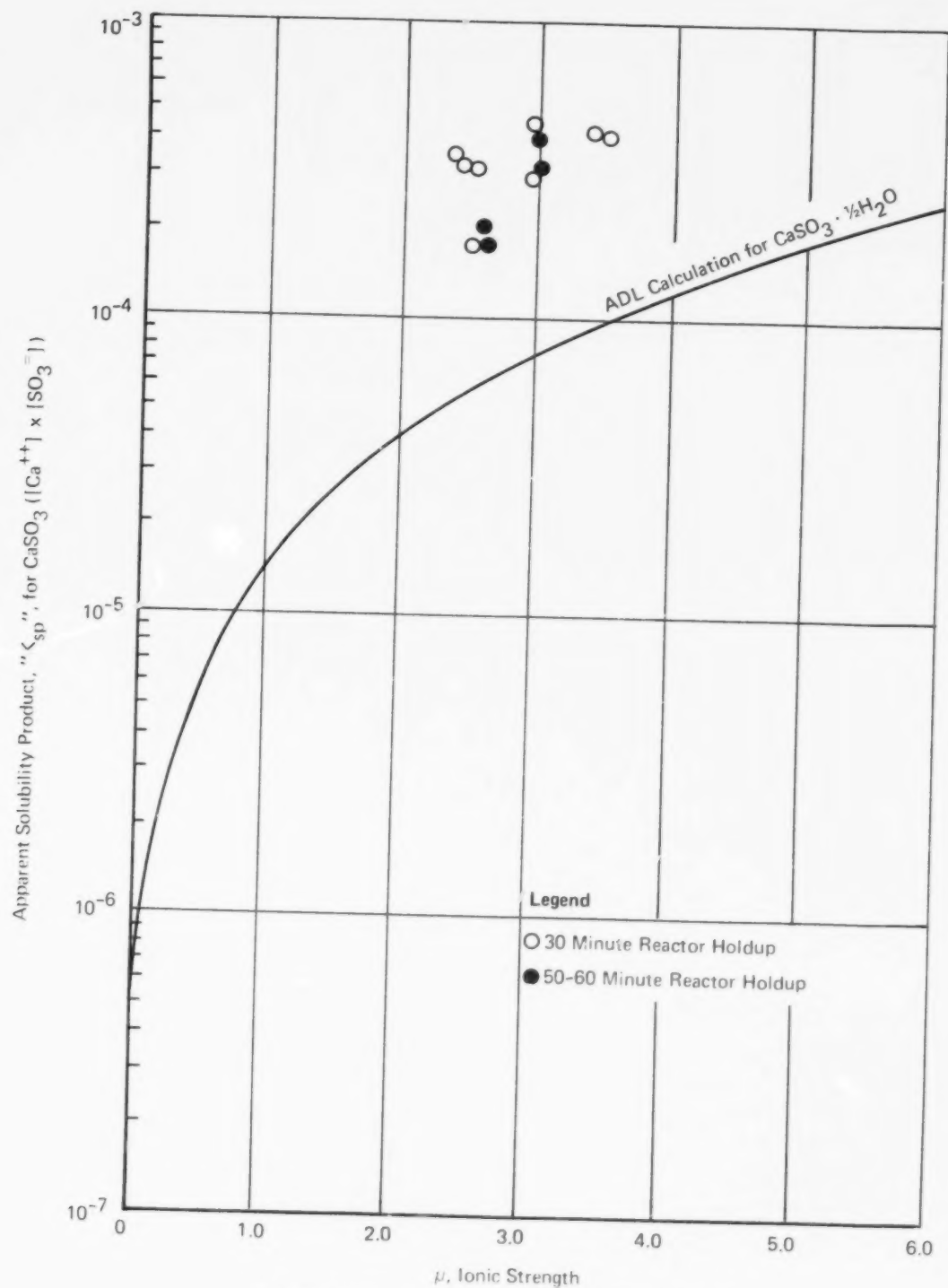


Figure III-5 Relation of Observed Apparent Solubility Products to Saturation Values For CaSO_3

There are two other explanations for this apparent supersaturation. First, the predicted curve may be slightly low. This curve was calculated based upon literature data for low ionic strengths using the method formulated by Kusik and Meissner for calculating activities at high ionic strengths. The method is only good to about 10 - 20% and any error in the data base would be translated directly into the predicted curve.

Second, the $\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$ solubility product curve may not apply, since the calcium sulfite/sulfate solids were apparently present as some type of mixed crystal as discussed in Section III. B. However, in no case did supersaturation to the extent it may have been present, cause any precipitation or scaling problems.

c. Regeneration of Active Sodium from Na_2SO_4 A careful analysis of the solids generated in the reactor showed that a significant amount of calcium sulfate precipitation was occurring over a wide range of soluble sulfate concentration. The amount of calcium sulfate precipitated was found to be strongly related to the soluble sulfate concentration. This sulfate dependence is shown in Figure III-5, in which observed mol ratios of $(\text{CaSO}_4/\text{CaSO}_x)$ in the reactor effluent solids, corrected for the small amount of occluded or entrained Na_2SO_4 , are plotted against sulfate concentration for TOS levels in the range 0.4 - 0.47M. The general relationship indicated by the dashed line agrees quite closely with that developed from the laboratory program for the same TOS range. The data suggests that there is a threshold level of ~0.05M sulfate at this TOS concentration above which precipitation of calcium sulfate occurs. This has not been confirmed and may simply be due to experimental and analytical error.

The results shown in Figure III-6 include data from runs with both the ADL reactor and the CSTR for reactor holdup times of 30, 45, and 60 minutes. In general, no significant effect of either reactor configuration or holdup time on the relative amount of sulfate precipitated could be discerned.

The results of a few runs made over a two-fold range of inlet TOS concentration, though, do suggest a dependence of sulfate precipitation on TOS. $(\text{CaSO}_4/\text{CaSO}_x)$ ratios obtained in three runs at 0.74 - 0.8M SO_4^{2-} , with TOS concentrations of 0.2, 0.4 and 0.52M are shown in Figure III-7. Although there is considerable data scatter, the observed decrease in sulfate precipitation for the two-fold plus increase in TOS is still significant. The dashed line shown in Figure III-7 is not meant to imply a straight line relationship between TOS and $(\text{CaSO}_4/\text{CaSO}_x)$; rather it is merely intended to be an indication of the trend in the data. A more definitive determination of the actual relationship between calcium sulfate precipitation and TOS concentration will require further study.

Laboratory studies of the stability of the CaSO_4 precipitated showed that significant redissolution of CaSO_4 solids (up to 30% of the CaSO_4 formed) may occur by continual intimate contacting of the solids with liquor containing appreciable levels of sulfite. It might be expected then that CaSO_4 redissolution together with additional CaSO_3 precipitation will occur in the dewatering system. No such redissolution could be substantiated in

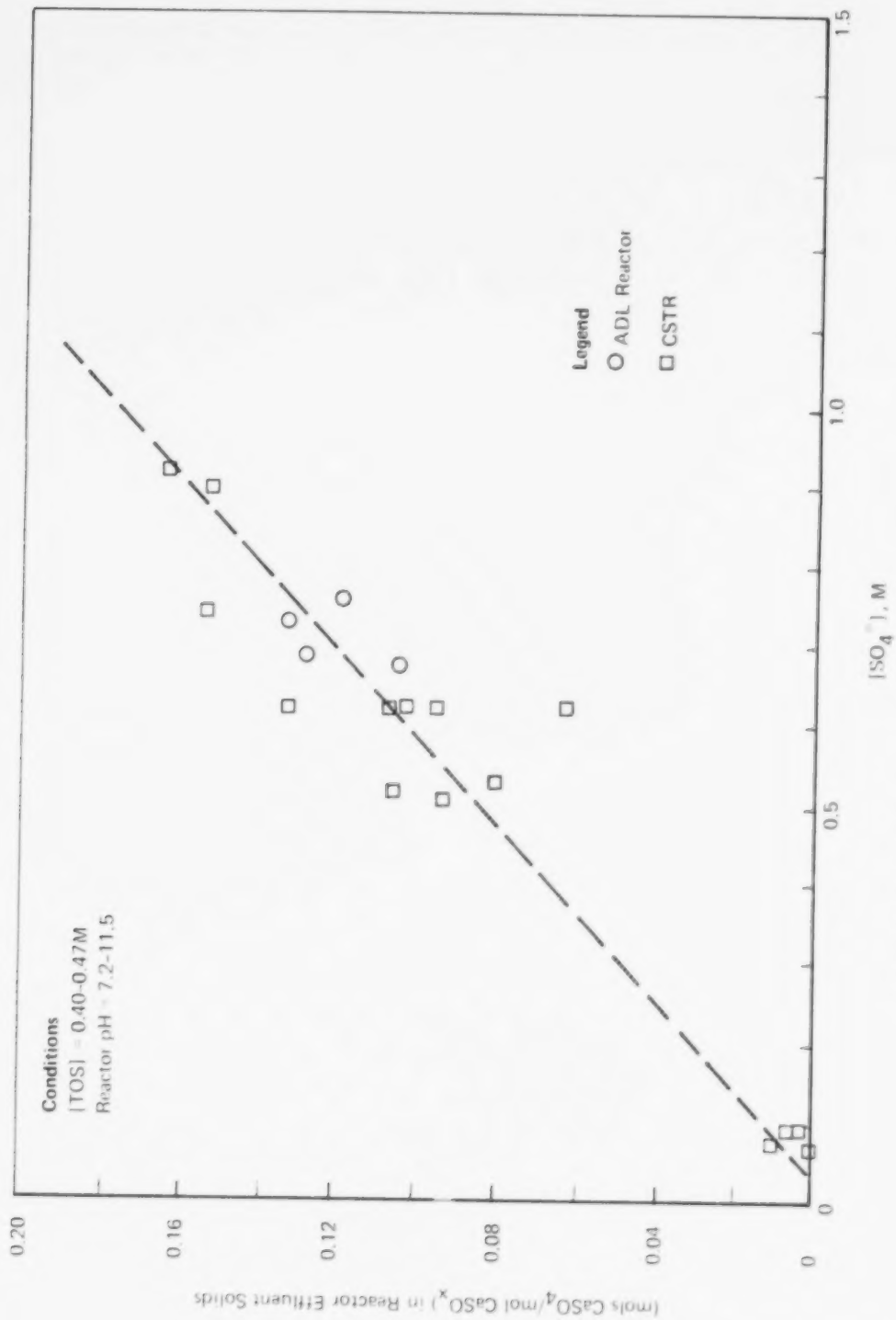


Figure III-6 ($\text{CaSO}_4/\text{CaSO}_x$) Ratio in Solids as a Function of Sulfate Concentration

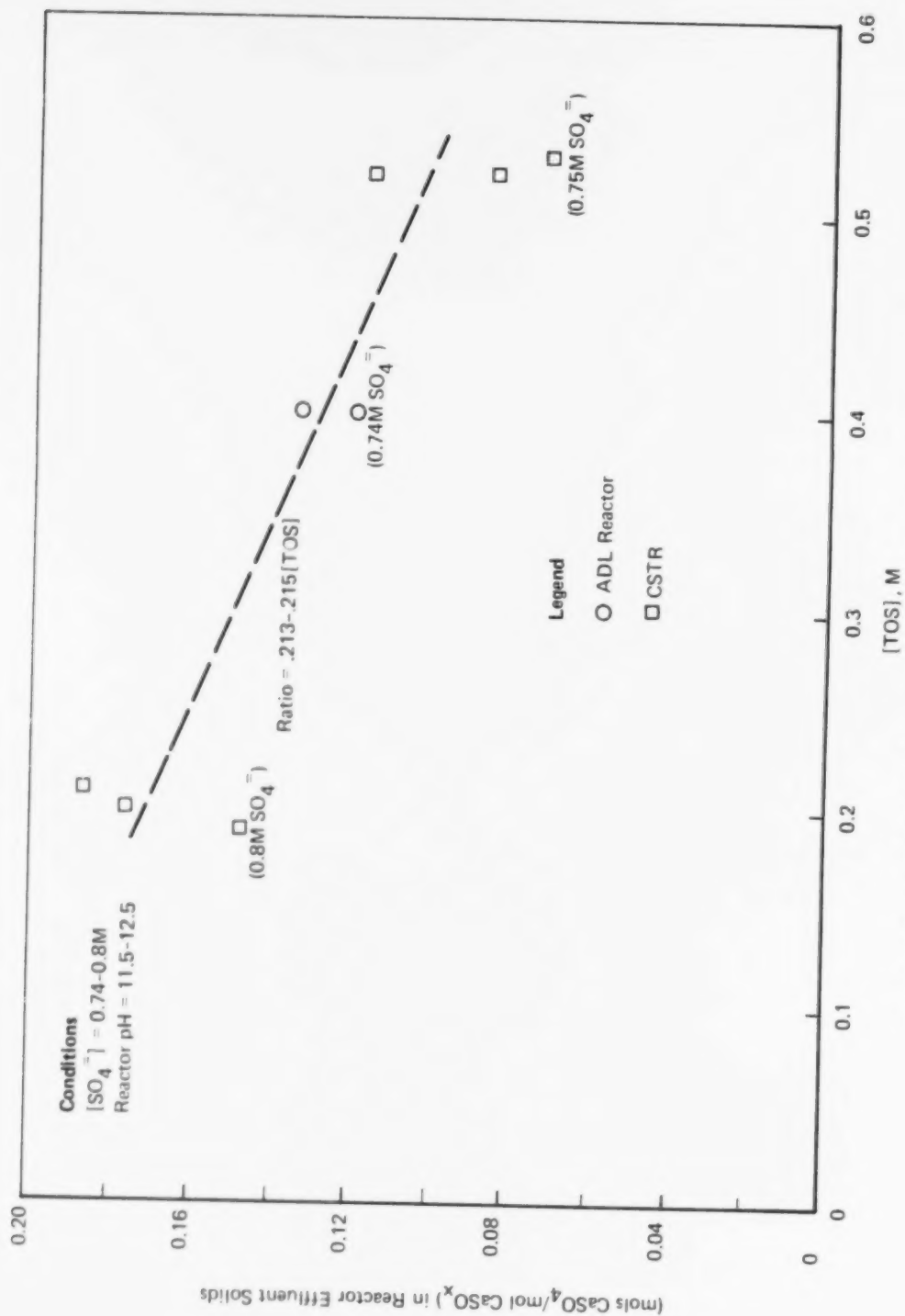


Figure III-7 $(\text{CaSO}_4/\text{CaSO}_4\text{x})$ Ratio in Solids as a Function of TOS Concentration

the pilot plant operations based upon a comparison of solids composition in the reactor effluent with that of the filter cake. The settler does not provide particularly good solids/liquid mixing as compared to holding the slurry in a well-mixed state for three hours, so the absence of redissolution is reasonable.

The amount of calcium sulfate precipitated in the system can be translated directly into the tolerable levels of sulfite oxidation in the process. In the range of sulfate concentration up to $1.0M SO_4^{2-}$ and active sodium levels of $0.2 - 0.4M Na^+$, it is possible to sustain an overall system oxidation rate equivalent to as much as 15 - 20% of the SO_2 removed without the need for a separate sulfate purge. The data suggest that the system could be operated at even higher sulfate concentrations with correspondingly higher sulfate precipitation capability. Such an approach of allowing sulfate levels to rise until sulfate precipitation balances sulfate formation however, will ultimately be limited by the solubilities of the sodium salts. Furthermore, precipitated calcium solids properties tend to deteriorate with increasing sulfate concentrations (as will be discussed). Therefore, the concentrated lime mode designed without special provisions for separate sulfate removal appears to be limited to applications where overall system oxidation is on the order of 20-25% or less of the SO_2 removed.

d. Dewatering Properties of Solids Although the performance of the ADL reactor and the CSTR were found to be essentially equivalent in terms of both calcium utilization and the regeneration of active sodium from soluble sulfate, there was a noticeable difference in the quality of the solids generated. The ADL reactor is apparently capable of successfully operating over a wider range of system conditions than the CSTR, at least with respect to the production of waste solids with acceptable settling and filtration properties.

CSTR Solids Properties The effects of reactor holdup and operating pH on the dewatering properties of the solids generated in the CSTR were initially studied in the preliminary open-loop CSTR reactor tests. Table III-2 gives the results of two sets of CSTR tests at $0.6M$ sulfate. The quality of the solids produced at a pH of 7.5 was good at both 30 minute and 60 minute reactor holdup times. As pH was increased to 12.0, the quality of the solids deteriorated as shown by the marked decrease in the solids settling rates and in the densities (wt% solids) of the settled solids. This pH effect was much more pronounced for the 60 minute holdup than for the 30 minute holdup; however, in neither case was the loss of solids properties irreversible. Reducing pH back to a low level resulted in a return to good solids. In light of the greater sensitivity of the 60 minute reactor holdup, the shorter 30 minute holdup provides a more reliable design basis for a CSTR.

The effect of sulfate concentration on solids properties over a range of pH for the 30 minute CSTR is summarized in Table III-3. Increasing sulfate levels results in poorer solids, an effect which is aggravated by high pH. The observed decrease in settling rates is indicative of smaller particle

Table III-2 EFFECT OF CSTR pH AND HOLD-UP TIME ON SOLIDS PROPERTIES --- CONCENTRATED LINE MOLE

General Reactor Conditions: Reactant Feed - Dry $\text{Ca}(\text{OH})_2$
 Reactor Temperature = 105 - 120°F
 Agitation Rate = 200 - 250 rpm
 Liquor Feed Composition - $[\text{TCS}] = 0.46\text{M}$
 $[\text{SO}_4^{--}] = 0.6\text{M}$
 pH = 5.8

Reactor Hold-up (min)	REACTOR EFFLUENT		SOLIDS SETTLING PROPERTIES	
	pH	Wt% Susp'd Solids	Initial Settling Rate (ft/min)	Wt% Insolubles in Settled Solids (2-3 hrs.)
60	7.2-7.4	2.5	0.16 - 0.23	26 - 36
	12.0	5.0	0.05	9
30	7.5	2.8	0.20 - 0.25	24
	11.9	4.7	0.15 - 0.20	15

Table III-3 EFFECTS ON SULFATE CONCENTRATION AND pH ON SETTLING PROPERTIES
OF SOLIDS PRODUCED ON A CSTR WITH A 30-MINUTE HOLDUP

General Reactor Conditions: Reactor Holdup = 30 minutes
 Reactant Feed - Dry $\text{Ca}(\text{OH})_2$
 Reactor Temperature = 105 - 120°F
 Agitation Rate = 200 - 250 rpm
 Liquor Feed Composition - $[\text{TOS}] = 0.45-0.55\text{M}$
 pH = 5.8

REACTOR EFFLUENT			SOLIDS SETTLING PROPERTIES	
Feed $[\text{SO}_4^{=}]$, M	pH	Approximate Wt. % Susp'd Solids	Initial Settling Rate, ft./min	Approximate Wt. % Insolubles in Settled Solids (2-3hrs.)
0.05	7.5	2.7	0.25 - 0.30	45
0.5	7.5-8.5	2.6	0.20 - 0.25	25
0.6	7.5	2.8	0.20 - 0.25	25
	11.9	4.7	0.15 - 0.20	15
0.75	12.0	4.4	0.05 - 0.10	15 - 20
	12.5	4.9	0.05 - 0.10	10 - 15
0.8	11-11.5	1.5	0.10 - 0.15	10
0.9-1.0 ^a	8-10	2.3	<0.05	5 - 15
	11.5-12.0	2.3	<0.05	5

(a) 45 minute reactor holdup

sizes; and the lower densities of the partially compacted solids implies a change in the form of the crystal agglomerates generated. Electron microscopy and optical studies of the solids produced confirms the difference in the precipitated crystals. The poorer settling solids are generally made up of smaller particle sizes with more needle-like crystal structures.

These differences in the settling properties are also reflected in the filterability of the solids. In general, low settling rates and low densities of settled solids are consistent with low solids content in the filter cake and poor cake wash efficiencies.

Comparative Solids Properties Table III-4 compares the settling and filtration characteristics for solids produced in the concentrated lime mode using both the ADL reactor and the CSTR. These results show that the density of the compacted solids, rather than the initial settling rate, is probably a better indication of the filterability and washability of the solids. This is reasonable since the density of the settled solids is simply a measure of how well particles compact under the force of gravity. Solids which do not compact well will not dewater easily, nor will they be easily washed.

As evidenced by the data presented in Table III-4, both reactors are capable of producing solids with good dewatering properties over a reasonable range of system operating conditions. In both systems, the solids properties deteriorate with increasing sulfate concentration and pH; however, the CSTR is more sensitive to these effects than the ADL reactor. Solids generated in the CSTR at high pH and sulfate concentration do not settle or filter as well as those obtained in the ADL reactor under equivalent conditions. This difference is shown by the plot of filter cake solids content vs. sulfate concentration in Figure III-8. The solid line represents filter data for the system operating with the ADL reactor which were obtained in work prior to this EPA program.

Washability of Solids The lowest levels of solubles obtained under continuous operation were 1.5 - 3.5 wt% solubles (dry basis). This was achieved in a run using slightly more than three displacement washes sprayed onto the cake using two full cone nozzles arranged in series. As indicated in Table III-4, the reduction in solubles is a function not only of the number of displacement washes but also of the quality of the solids. As the quality deteriorates, the moisture content of the cake increases and the wash efficiency (and percent reduction in solubles) decreases.

The practical minimum solubles levels which can be achieved was determined in a special set of tests conducted with the pilot plant filter. The solids used were produced in the ADL reactor at a 0.6M sulfate concentration. The results of these tests demonstrated that the cake can be realistically washed to a level of about 1.5 wt% soluble solids (dry basis), if desired. The reduction in soluble solids attained as a function of the number of displacement washes used is shown in Figure III-9. Here a reduction of 90% in the soluble solids is roughly equivalent to 1.2 wt% soluble solids (dry basis) for a cake containing 50 wt% insoluble solids. This level of sodium

TABLE III—(Continued) SOLUBLE AND INSOLUBLE IN 10% ETHANOL, 10% 80% 90%

Solvent system	Solvent conditions	Swelling Properties			Filterability			Filter Losses, Insoluble			Total Efficiency, %
		ΔV , ml/gm. dry ash	ΔV , ml/gm. dry ash	ΔV , ml/gm. dry ash	ΔV , ml/gm. dry ash	ΔV , ml/gm. dry ash	ΔV , ml/gm. dry ash	No. of particles per ml. of filtrate	No. of particles per ml. of filtrate	No. of particles per ml. of filtrate	
A-1	10 min.	0.7	0.11	0.16	0.16	0.16	0.16	1.6	1.6	1.6	90-100
	60 min.	0.8	11.0-12.0	0.13	0.13	0.13	0.13	1.6	1.6	1.6	100-100
C-18	30 min.	0.3	7.3-8.3	0.22	0.22	0.22	0.22	1.3-1.5	1.3-1.5	1.3-1.5	80-100
	45 min.	0.3	11-12.5	0.12	0.12	0.12	0.12	1.3	1.3	1.3	100-100
	45 min.	0.3-1.0	8.0	0.09	0.09	0.09	0.09	1.3	1.3	1.3	100-100

a. No. of particles per gram = $\frac{\text{No. of particles}}{\text{No. of water in cake}}$

b. Type of spray: (1) jet—single, full-cone spray; (2) jet—two, full-cone sprays in series

c. Bath efficiency is calculated by dividing the observed reduction in soluble by the reduction expected assuming dilution of the cake liquor. For example, in the first test shown, the expected reduction in soluble based upon dilution is $1 - (1/2.5)^{1/2} = 80\%$, and the observed reduction is $55 - 75\%$. Therefore, the efficiency is $80/125 = 64\%$.

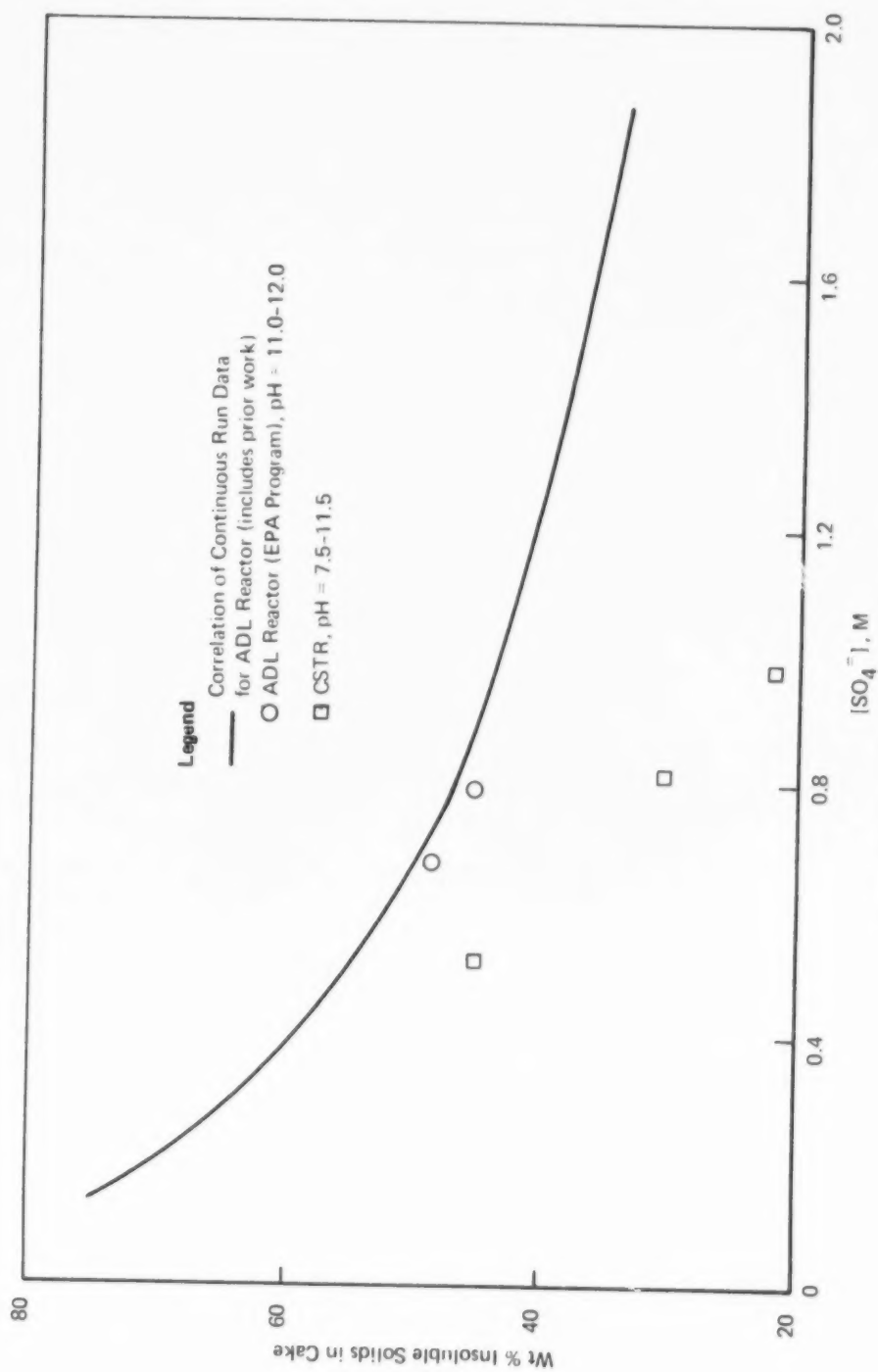


Figure III-8 Average Wt % Insolubles in Filter Cake Versus [SO₄²⁻] — Concentrated Lime Mode

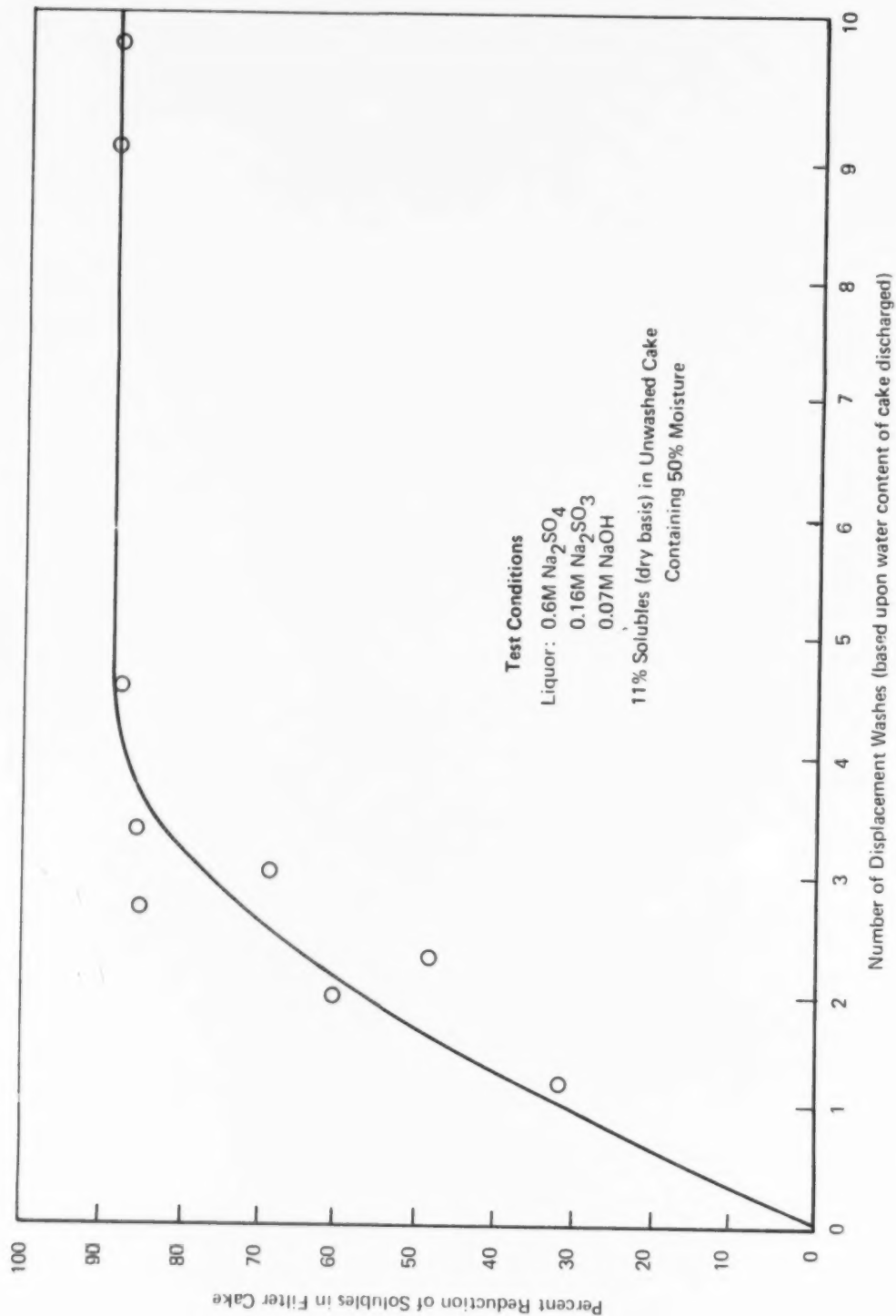


Figure III-9 Reduction of Solubles in Filter Cake Versus Wash Ratio
(Two Sprays in Series)

salts (almost all of which are sodium sulfate) would leave the system at any reasonable washing rate. This confirms similar data obtained in the laboratory. There is apparently a low but significant quantity of soluble salts, on the order of 1 - 1.5 wt% solubles, occluded in the cake that is not easily leached.

The maximum cake washing capability for a medium or high sulfur coal application would be roughly equivalent to two to three displacement washes (0.4 gpm/Mw for a cake with 50 wt% solids). The number of displacement washes will be limited by the amount of makeup water available for washing as determined by the process water balance and possibly by the operation of the filter. Assuming cake wash behavior similar to that seen in the ADL pilot plant for two serial spray washes, 75 - 85% reduction in soluble solids concentration in the cake liquor can be expected for this level of washing. With a 0.6 - 0.7M sulfate concentration in the system liquor, the resultant solubles in the cake would then be 2 - 3% (dry cake basis) for a cake containing 50 wt% insoluble solids.

The exact amount of solubles will depend upon a large number of factors including dissolved solids levels, the type of wash used (spray configuration), the operating conditions for the filter, and especially the solids properties. The degree of washing that is desirable will also depend in large part on the presence of impurities -- the rate at which they enter the system and the tolerable steady state impurity levels within the system.

e. Reactor System Oxidation A significant amount of oxidation was found to occur in both reactor systems operated open to the atmosphere. In the initial open- and closed-loop continuous reactor tests, oxidation rates ranged from 0.01 to 0.30 mols TOS/liter-min (equivalent to 10 - 280 ppm of SO₂ in the pilot plant system). The rate of oxidation was found to be essentially independent of sulfite concentration in this concentrated regime; but it was strongly dependent upon the degree of agitation and the exposed surface area of the vessel indicating that the oxidation is primarily a function of the amount of gas-liquid contacting.

Subsequent batch and continuous runs showed that the oxidation in the reactor system could be minimized by controlling agitation and modifying the design of the vessel; and that it can be virtually eliminated by sealing the reactor system. Based upon these results, it would normally be expected that oxidation rates in an unsealed reactor system would be in a range equivalent to 50 ppm SO₂.

3. Summary of Overall System Operation

Four closed-loop continuous runs were made in the pilot plant in the concentrated active sodium mode using lime for absorbent regeneration. A range of different conditions were examined including SO₂ inlet concentration (2600 - 2800 ppm and 850 ppm), sulfate concentration, reactor system design (ADL vs. CSTR) and the use of both Na₂SO₄ and Na₂CO₃ for sodium makeup. For the most part, these runs were intended to confirm and extend open-loop data and to test the system operation under a variety of conditions.

In each run, the system was primed to predetermined concentrations and then operated under a specified set of conditions for at least three days (four to eight system holdup times). Although steady state operation was not achieved in all runs, sufficient data were collected to characterize the operation of each section for the given set of conditions.

Under all conditions tested, the system operation was found to be quite stable. No difficulties were encountered in attaining SO_2 removal efficiencies greater than 90%. Lime utilizations exceeded 95% throughout all runs and soluble calcium levels in the regenerated absorbent liquor returning to the scrubber were consistently less than 100 ppm. No scaling occurred in the system during these runs and no increase in tray pressure drop was observed.

The amount of oxidation experienced in the system, the properties of the solids generated (including the cake properties) and the rate of regeneration of active sodium from sodium sulfate differed in each run depending upon the particular set of conditions being tested. However, two of the runs do provide a basis for analyzing the operation of systems incorporating the CSTR and the ADL reactor. The primary system operating conditions and key operating results for these runs are summarized in Table III-5.

Run 404 was the last continuous run made in the concentrated lime mode. In this run, using a new CSTR reactor, oxidation was reduced in the reactor and maximum cake wash was used (three displacement washes). Run 402, using the ADL system, was conducted prior to reactor redesign for oxidation control and incorporated only minimum washing of the filter cake (one displacement wash). As a consequence, reactor oxidation and cake solubles content were higher in the earlier run 402 than 404.

It should be noted that not all process sections were operating under optimal conditions in these runs, although overall they represented system operating levels which might be achieved under steady state conditions.

Both runs 402 and 404 were made at high SO_2 inlet levels and active sodium concentrations of about 0.45. In run 402, incorporating the ADL reactor, there was a slight decrease in active sodium throughout the run due to the low cake wash rate and the relatively high oxidation rate in the reactor system. Na_2SO_4 was also used to make up sodium value even though the sulfate formation rate within the system was slightly higher than the sulfate precipitation.

In run 404, the CSTR run, the system was operated at a much lower sulfate level than run 402 to ensure the production of solids with good dewatering properties. Although there was a lower rate of regeneration of Na_2SO_4 to active sodium due to the lower sulfate level, the use of a reactor vessel with a smaller surface area than that used in run 402, which minimized oxidation in the reactor, coupled with the slightly higher SO_2 inlet level, enabled the system to essentially sustain a rate of sulfate precipitation equivalent to the sulfate formation rate.

Table III-5 SUMMARY OF CLOSED-LOOP RUNS 402 AND 404 -- CONCENTRATED LIME MODE

Test No.	402	404
<u>Operating Conditions</u>		
Inlet Gas SO ₂ Level, ppm	2600	2800
Scrubber Bleed: pH	5.8-5.9	5.8-6.0
[TCS], M	0.45	0.4
[SC ₄ ⁼], M	0.85-0.8	0.5
Reactor System: Type	ADL	CSTR-30 min. holdup
pH	11.5-12	7.5-8.5
Sodium Makeup Form	Na ₂ SO ₄	Na ₂ CO ₃
<u>Key Operating Results</u>		
Scrubber Feed Stoichiometry	1.1	1.1-1.2
SO ₂ Removal Efficiency (% of inlet SO ₂)	91%	93-95%
Calcium Utilization (% of available Ca(OH) ₂)	96%	98+
Calcium Feed Ratios:		
mols Ca(OH) ₂ /mol/ΔSO ₂	1.10	0.92(0.96 calcium)
[Ca ⁺⁺] in Scrubber Feed, ppm	60	30-70
System Oxidation Rates:		
Scrubber Circuit (equiv. ppm of SO ₂)	200	210
Regeneration Reactor (equiv. ppm of SO ₂)	240	30
Sulfate Precipitation (equiv. ppm of SO ₂)	330	230
Filter Cake Composition:		
wt% insoluble solids	45wt%	45wt%
wt% soluble solids (avg.)	6wt%	2wt%
No. of Displacement washes	1	3

Figures III-10 and III-11 summarize the key operational characteristics for runs 402 and 404 in terms of schematic block process diagrams. These diagrams express the various system inputs and outputs in terms of the SO_2 absorbed. For each of the three process sections (gas scrubbing, regeneration reactor, and dewatering) the observed rates of sulfite oxidation and/or regeneration of active sodium from sodium sulfate are included, also expressed as a function of SO_2 absorption. Oxidation (and regeneration) rates in the scrubber and reactor systems are determined from material balances around each of these process sections individually. Oxidation rates in the dewatering system, on the other hand, were extracted from overall system material balances over a period of at least 24 hours, taking into account changes in inventory within the system. The direct determination of sulfite oxidation rates across the dewatering system were difficult to accurately estimate since flows were not constant, and wash water and vacuum pump water were being continually added to the system through the filter. The best direct estimates of oxidation across the filter are 0.02 - 0.04 gmols TOS/min; and across the entire dewatering system is about 0.05 gmols TOS/min. This overall rate is roughly equivalent to the oxidation of 50 ppm SO_2 in the scrubber and agrees closely with the rates calculated from overall balances.

The composition of the cakes shown in Figures III-10 and III-11 are given in terms of wt% on a total dry basis. The value shown for CaSO_4 content in the cake, therefore, differs slightly from the rate of precipitation shown in the regeneration system block.

As can be seen in the composite for run 402, sodium makeup is roughly equivalent to sodium losses in the cake; however, the total sulfate precipitated and lost in the solubles fraction of the cake amounts to only about 20% of the SO_2 absorbed while the overall system oxidation plus sulfate addition totaled 27% of the SO_2 absorbed. There was, therefore, a slow decrease in active sodium in the system and a corresponding increase in sulfate. This would have continued until the level of sulfate rose to a point where it would effect high enough sulfate precipitation rates to provide a balance in the system.

It should be noted that in run 402, the total calcium feed to the system amounted to 110% of the SO_2 absorbed. This feed rate was simply high, and does not reflect a low utilization of lime. This amount of extra lime is not needed for the reaction with the sulfur added to the system as Na_2SO_4 , since this sulfur would normally leave in the cake associated with sodium that is being discharged.

Composite Diagram of Pilot Plant Operations Figure III-12 summarizes the operational characteristics of the various pilot plant process sections in the concentrated lime mode as determined in the various open- and closed-loop runs. This block diagram is analogous to those for run 402 and 404, and shows the expected steady state operation of the dual alkali system for a 2500 ppm inlet SO_2 loading with a total system oxidation rate of 13%. The composite is based upon operation with the ADL reactor. For a comparable operation using the CSTR, the solids content

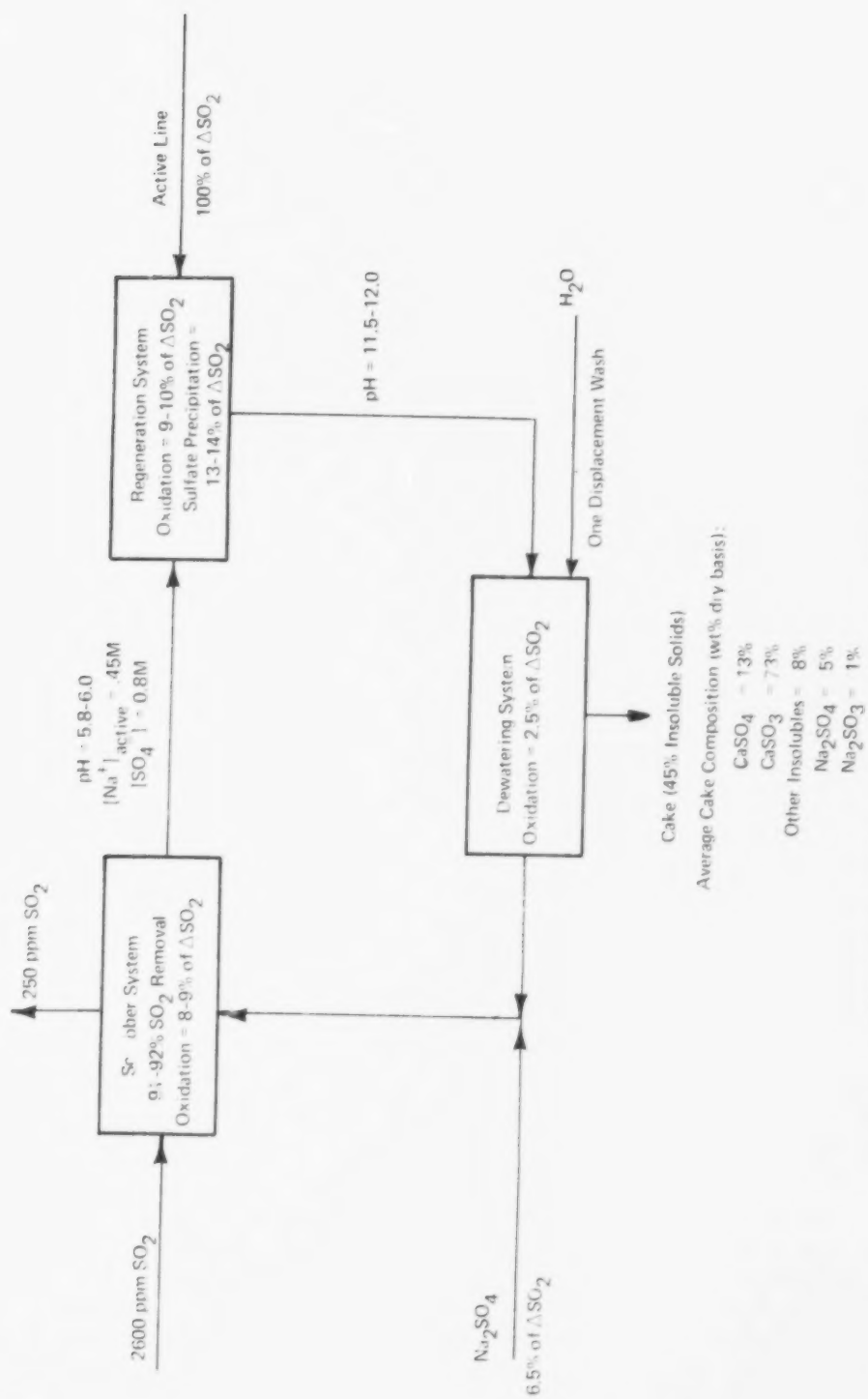


Figure III-10 Continuous Run 402 — ADL Reactor (Na_2SO_4 Makeup)

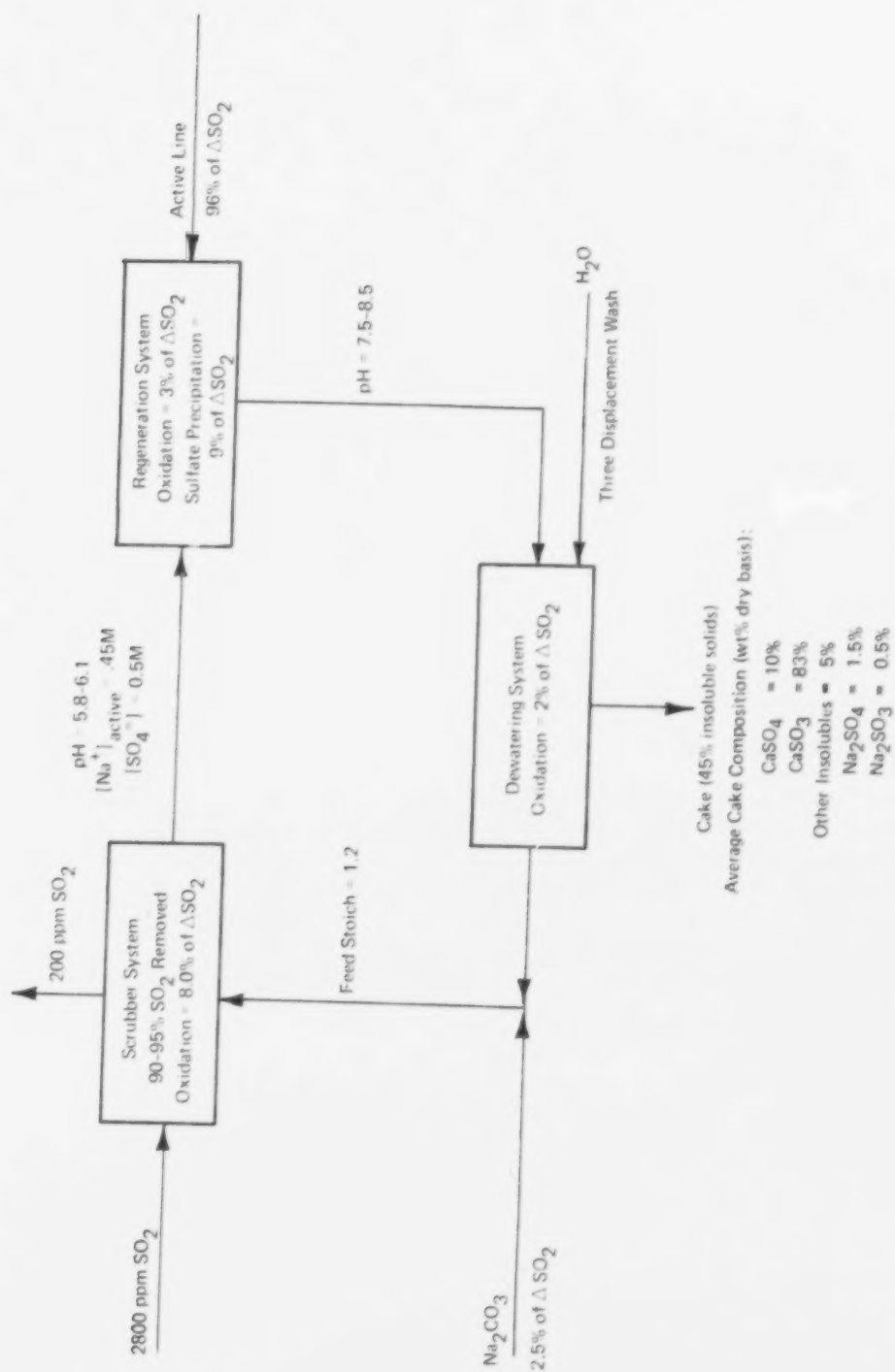


Figure III-11 Continuous Run 404-CSTR Reactor (Na_2CO_3 Makeup)

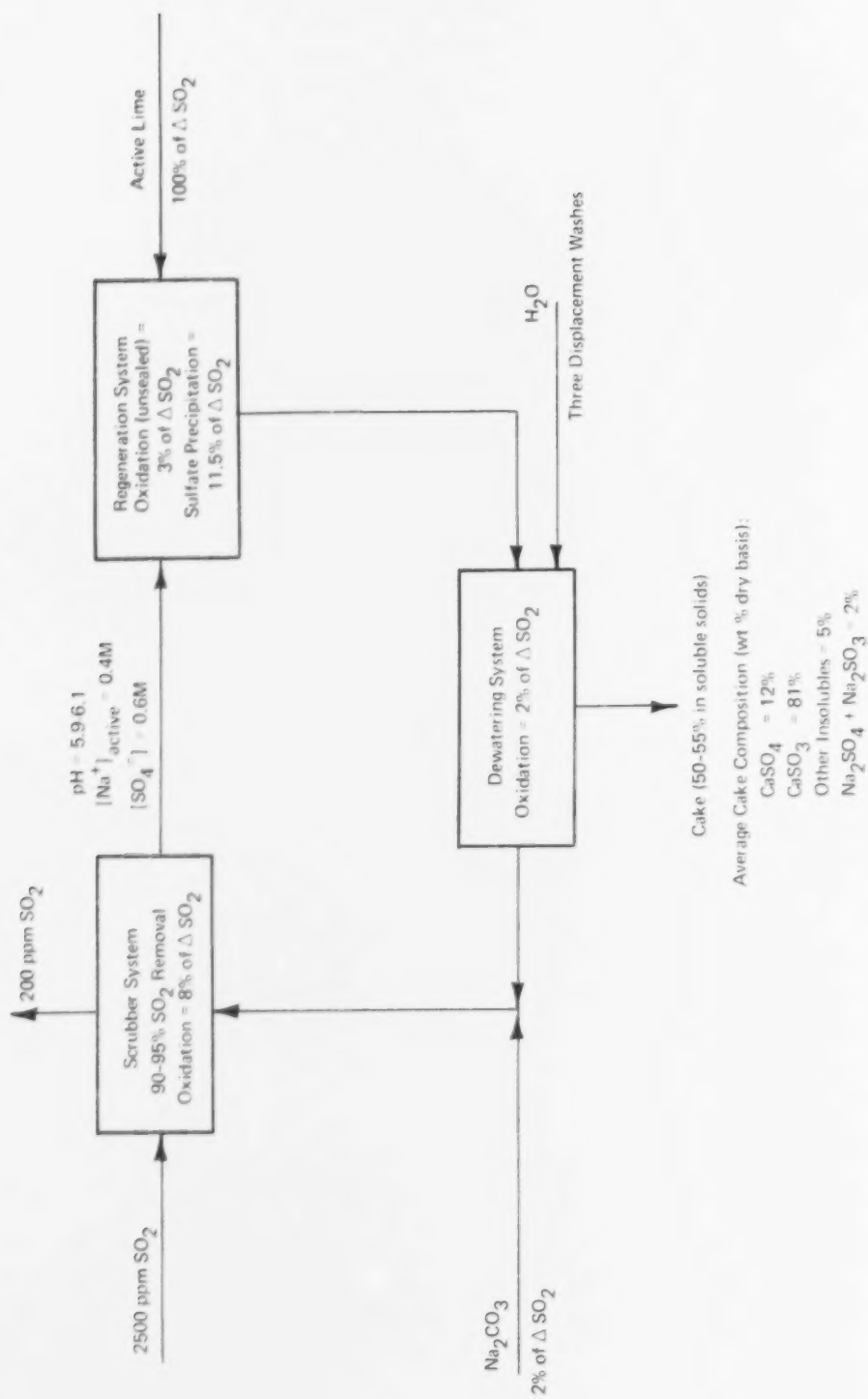


Figure III-12 Composite Pilot Plant Operations For 13% Overall Oxidation in System

of the cake would be expected to be 40 - 45% insoluble solids. These solids contents are estimated from pilot plant data and for larger scale systems where the dewatering equipment can be operated with better control, filter cakes with up to 5 - 10% higher solids levels would be expected.

4. Future Pilot Plant Operations

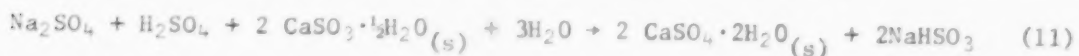
Future pilot plant work in the concentrated lime mode will involve a number of four to five week closed-loop continuous runs using the CSTR and/or ADL reactor system. These runs will be aimed at demonstrating the overall system performance and specific emphasis placed on process reliability and stability and minimizing soluble solids losses in the cake (within practical limitations of wash capability).

IV. SULFURIC ACID TREATMENT -- CONCENTRATED MODE

A. LABORATORY RESULTS

1. Introduction

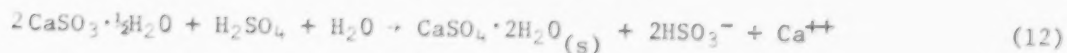
The inherent sulfate precipitation which was observed during lime regeneration of concentrated sodium sulfite/bisulfite/sulfate solutions shows promise of being able to control sodium sulfate at manageable levels when system oxidation rates are in the range of 10 - 20%. An alternate method, capable of dealing with significantly higher oxidation rates, involves the acidification of the process liquor with sulfuric acid in the presence of solid calcium sulfite to precipitate dissolved sulfate as gypsum. The overall reaction involved in this treatment process is,



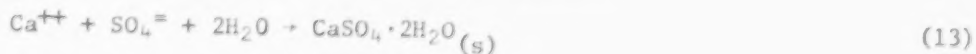
thus, although additional sulfate is introduced as sulfuric acid, both that sulfate and the Na_2SO_4 should be precipitated as gypsum. On the basis of the 1:1 relationship between sulfate (sodium) to be precipitated and sulfuric acid required, a reaction efficiency (theoretically 100% based on equation 11) can be defined by expressing the moles of Na SO removed from the process stream as a percent of the moles of H_2SO_4 added:

$$\text{Efficiency} = \left[\frac{\text{mol Na}_2\text{SO}_4 \text{ removed}}{\text{mol H}_2\text{SO}_4 \text{ fed}} \right] \times 100\%$$

The primary acid consuming reaction is the dissolution of $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$ by converting it to the soluble calcium bisulfite



The dissolved calcium ion produced in this reaction can then react with the sodium sulfate to be precipitated to form gypsum:



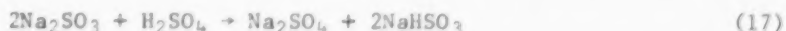
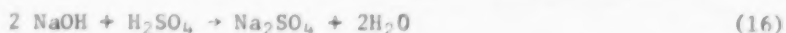
In addition to the primary reaction discussed above, other equilibria would likely be involved and would result in reaction efficiencies of less than 100%. At the acidic conditions required to dissolve calcium sulfite ($\text{pH} \approx 2 - 3$), measureable amounts of sulfurous acid and possibly bisulfate

ion could exist in solution due to the acid equilibrium reactions:

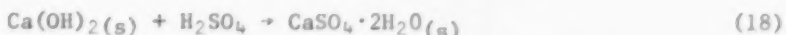


In practice, the streams to the acid treatment could contain other sulfites, carbonates and possibly hydroxide. In such cases, sulfuric acid consumption would be increased and reaction efficiency reduced.

The most acidic process stream normally available from which sulfate could be precipitated -- scrubber effluent -- normally contains on the order of 0.05 M Na_2SO_3 . Even well washed filter cake could contain calcium carbonate or unreacted lime. Depending upon washing efficiency, significant amounts of Na_2SO_3 and NaOH could also be present. If the calcium sulfite were obtained as a thickened slurry from the regeneration section -- attractive because of the ease of obtaining it and transferring it to the acidification process -- significantly larger amounts of NaOH and Na_2SO_3 would be present. The NaOH and Na_2SO_3 will react with H_2SO_4 according to the reactions



consuming H_2SO_4 and producing additional soluble sulfate in solution. Any lime or limestone present in the calcium sulfite feed will react with sulfuric acid



These reactions will consume additional acid but will not add additional sulfate to the solution because of the accompanying precipitation of gypsum.

As a point of departure for the subsequent laboratory studies, a very approximate estimate was made of the pH to which the solution would need to be taken for the regeneration reaction to proceed, based upon very dilute solution solubility product and acid dissociation constants. For the precipitation of gypsum to occur, its solubility product in water

$$K_{\text{sp}} = [\text{Ca}^{++}] [\text{SO}_4^{=2-}] \approx 2 \times 10^{-6}$$

would need to be exceeded. If $[\text{SO}_4^{=2-}]$ were 1.0M initially, $[\text{Ca}^{++}]$ would need to exceed 2×10^{-6} M for precipitation to occur. For the precipitation of gypsum to continue until $[\text{SO}_4^{=2-}] = 0.1$ M, the calcium ion concentration in solution would need to rise further to about 2×10^{-5} M.

To achieve calcium levels of this magnitude in the presence of the very insoluble calcium sulfite, $[\text{SO}_3^{=2-}]$ would need to be held to a very low level by converting it to bisulfite ion. Considering the solubility product of

calcium sulfite,

$$K_{sp} = [Ca^{++}] [SO_3^{--}] = 10^{-7}$$

a calcium ion concentration of about 10^{-3} would require that the level of sulfite ion be held to about 10^{-4} M. Now, to achieve that sulfite ion level in a solution containing sulfite/bisulfite at about 0.5 M, the second ionization constant of sulfurous acid

$$K = \frac{[SO_3^{--}] [H^+]}{[HSO_3^-]} = 10^{-7}$$

would indicate that $[H^+] = 5 \times 10^{-3}$, or a pH in the vicinity of 3, would be required.

2. Experimental Results and Discussion

Prior to studying the acidification reaction in the laboratory in a continuous reactor, a number of batch experiments (in which H_2SO_4 was added to $Na_2SO_4/CaSO_3$ slurries) were conducted in a sealed reactor to better define the operating pH range and, more importantly, to establish the SO_2 partial pressure levels generated within the reactor. The $Na_2SO_4/CaSO_3$ slurries were charged into the reactor, and the headspace above the liquid was purged briefly with SO_2 to eliminate air before sealing the reactor. Sulfuric acid was then introduced into the sealed system from a pressure-equalized addition funnel while monitoring the pressure within the reactor.

The batch experiments indicated that the operating pH would, in fact, be in the range of 2.5 - 3.0. Throughout the course of sulfuric acid addition, the pressure within the reactor indicated that the partial pressure of SO_2 was less than 1 atmosphere, and a sealed reactor capable of withstanding a positive internal pressure would not be required for the reaction. The reactor could be operated at ambient pressure vented to the atmosphere as long as the reactor vent tube was of a sufficiently narrow bore to minimize mixing between the internal headspace gas and the outside air.

On the basis of the background information obtained from batch experiments, the effects of a number of experimental variables on the regeneration reaction were studied in the laboratory using the CSTR described in Section II-A. Since it was the only material available in quantity at that time, the calcium sulfite used in these studies was unwashed calcium sulfite filter cake produced in an earlier ADL pilot plant run in which concentrated sodium bisulfite/sulfite/sulfate solutions were regenerated with lime. This material contained relatively large amounts of calcium carbonate and calcium sulfate along with lesser amounts of sodium salts as the following composition indicates:

<u>Substance</u>	<u>Mol Percent</u>
CaSO ₃	69.2
CaCO ₃	15.4
CaSO ₄	10.5
Na ₂ SO ₄	3.3
NaOH	1.0
Na ₂ SO ₃	0.5

This composition is based upon chemical analyses and material balances for the pilot plant runs in which the material was produced. This filter cake was slurried with a simulated acidic scrubber effluent solution containing, in all cases, Na₂SO₃, 0.05 M; NaHSO₃, 0.40 M; and Na₂SO₄ at either 0.75 M or 0.37 M. This slurry comprised one stream to the CSTR and either 4.5 M or 9.0 M H₂SO₄ was fed as the other reactant.

Both phases of the reactor effluent slurry were analyzed in detail. In the liquid phase, in addition to measuring total soluble calcium ion and sulfate ion, measurements of Total Oxidizable Sulfur [TOS] and Total Acidity [H⁺]_{tot} were made by titration. The concentration of TOS is the total concentration of all sulfur (IV) species present, and the total acidity [H⁺]_{tot} is the sum of all acidic protons (titratable to pH = 9) in the solution. These relationships may be expressed as follows:

$$[\text{TOS}] = [\text{H}_2\text{SO}_3] + [\text{HSO}_3^-] + [\text{SO}_3^{=}]$$

$$[\text{H}^+]_{\text{tot}} = [\text{HSO}_3^-] + 2[\text{H}_2\text{SO}_3] + [\text{HSO}_4^-] + [\text{H}^+]$$

By subtraction, the "excess acidity", ([H⁺]_{tot} - [TOS]) equals

$$[\text{H}^+]_{\text{tot}} - [\text{TOS}] = [\text{H}_2\text{SO}_3] + [\text{HSO}_4^-] + [\text{H}^+] - [\text{SO}_3^{=}]$$

However, under the range of conditions over which the reaction was studied (see Table IV-1), H₂SO₃ and HSO₃⁻ are the predominant species present. At pH 2 - 3, [SO₃⁼] is virtually zero and [H⁺] is less than 0.01 M.

The dilute solution value for the second ionization of sulfuric acid, pK₂, is about 2; however, dissociation of HSO₄⁻ increases as ionic strength is increased. In the CSTR experiments performed, the ionic strength was in two general regions -- 2.25 (experiments 6 - 7) and 1.4 (experiments 9 - 10). Marshall and Jones⁶ report values of pK₂ for H₂SO₄ of 0.82 and 1.1, respectively, at those ionic strengths. At the conditions studied here, then, HSO₄⁻ exists only in small concentrations -- the concentration of HSO₄⁻ is at most 2% of the H₂SO₃ concentration in the solution. The two measurements [H⁺]_{tot} and [TOS], allow one to determine the amounts of H₂SO₃ and HSO₃⁻ present in the reaction effluent.

The results obtained from the CSTR studies are shown in Table IV-1. In the three series-6 experiments, the regeneration of 0.75M Na_2SO_4 solution was studied as a function of pH ranging from about 2.35 to 3.15 with reactor residence time and temperature held constant. In series 7, temperature and residence time were varied. In the final three experiments, 0.37M Na_2SO_4 was studied as a function of pH.

The effect on the regeneration reaction of reducing the pH from 3.15 to 2.9 was an increased TOS concentration as additional calcium sulfite dissolved at the lower pH. In turn, the concentration of $\text{SO}_4^{=}$ decreased as the soluble calcium which was produced reacted with $\text{SO}_4^{=}$ to precipitate gypsum. As indicated by the lower $\text{SO}_4^{=}$, the "effective Na_2SO_4 removal" from the simulated scrubber stream ($\Delta\text{Na}_2\text{SO}_4/\text{Na}_2\text{SO}_4$) was higher at the lower pH (this ratio is corrected for all sulfate removed which entered as H_2SO_4 or in the calcium sulfite filter cake).

As the pH was decreased to 2.9, the concentration of TOS increased by 0.1M but the total acidity increased by 0.17M as a result of the higher concentration of H_2SO_3 present at the lower pH. The ratio of $\text{H}_2\text{SO}_3/\text{HSO}_3^-$ had risen from 0.09 to 0.15 as shown in the table. The formation of H_2SO_3 consumed H_2SO_4 and should have reduced the efficiency of H_2SO_4 utilization. Thus the increased effective sulfate removal at pH 2.9 was compensated for by the decrease in efficiency of H_2SO_4 utilization to produce an essentially constant overall "observed efficiency" of H_2SO_4 utilization -- $\Delta\text{Na}_2\text{SO}_4/\text{total } \text{H}_2\text{SO}_4 \text{ fed}$.

Since the calcium sulfite cake which was used contained a large amount of CaCO_3 as well as other impurities which would consume H_2SO_4 , the effects of the non-calcium sulfite species were taken into account and a "corrected efficiency", which should have been realized if pure calcium sulfite had been fed, was computed. As shown in the table, H_2SO_4 utilization efficiency would have been in the range of 0.6 to 0.7.

In experiment 6-c, the pH was further lowered to 2.35; but, upon analysis of the reaction products, it was found that no solid calcium sulfite remained in the effluent slurry. The results of this experiment show the important consequence of operating "calcium sulfite limited" (less than the stoichiometric amount of calcium sulfite); both the effective Na_2SO_4 regeneration and H_2SO_4 utilization dropped significantly. In experiments 6-2 and 6-b, about 20% and 10%, respectively, of the calcium sulfite fed remained in the reaction solids.

Decreasing the reactor residence time to 15 minutes and raising the reaction temperature from 34 to 51°C (series 7) did not produce significant changes in effective sulfate regeneration or reaction efficiency.

When the concentration of Na_2SO_4 was reduced to 0.37M in experiment 9, an effective Na_2SO_4 removal of 0.66 was observed at pH = 2.8, a value slightly higher than in the corresponding experiment at the higher sulfate concentration (7-a); however, both the observed and corrected H_2SO_4

Table IV-1 SUMMARY OF LABORATORY CONTINUOUS REACTOR SULFURIC ACID TREATMENT EXPERIMENTS

Expt. ^{a, b}	Experimental Variables				Effluent Concentrations (M)					Efficiency		
	[Na ₂ SO ₃]	Residence Time (min)	Temp (°C)	pH	[TOS]	$[H^+][SO_3^{=}]$	[Ca ⁺⁺]	$\frac{[Ca^{++}][SO_3^{=}]}{(\times 10^5)}$	$\frac{[H_2SO_4]}{[HSO_3^-]}$	Effective Na ₂ SO ₃ Removal ^c	$\frac{\text{observed}^d}{\text{corrected}^e}$	
6-a	0.75	30	33	2.1	1.39 _e	1.51	0.35 _e	0.024	8.5	0.090	0.49	0.38
6-b	0.75	30	34	2.9	1.49	1.68	0.29	0.026	7.5	0.15	0.57	0.38
6-c	0.75	30	34	2.3	1.46	1.98	0.36	0.014	5.0	0.55	0.40	0.25
7-a	0.75	15	34	2.8	1.52 _e	1.68	0.27	0.034	9.2	0.11	0.61	0.42
7-b	0.75	30	34	2.8 _e	1.59 _e	1.75	0.25	0.030	7.5	0.11	0.63	0.40
7-c	0.75	15	51	2.8	1.56	1.72	0.25 _e	0.032	8.2	0.11	0.62	0.37
9	0.37	15	33	2.8	1.13 _e	1.29	0.14	0.041 _e	5.8	0.16	0.66	0.33
10-a	0.37	15	33	2.7	1.41 _e	1.60 _e	0.073 _e	0.069	5.1	0.16	0.80	0.29
10-b	0.37	15	33	2.2	1.56	1.84	0.055	0.107	5.9	0.22	0.85	0.22
												0.35

^a Solution to be regenerated simulated scrubber effluent:[NaHSO₃] = 0.4 M[Na₂SO₃] = 0.05 M[Na₂SO₄] = as indicated

pH = 5.4 - 5.6

^b Calcium sulfite feed maintained at 110 - 120% of stoichiometric except:Expt. 6-c was deficient; no CaSO₃· $\frac{1}{2}$ H₂O in effluent

Expt. 10-a, 10-b fed at 200% of stoichiometric

^c Mole fraction of entering Na₂SO₃ actually removed.^d (Moles Na₂SO₃ in - moles soluble SO₃⁼ out)/moles H₂SO₄ charged.^e Observed efficiency corrected for effect of alkaline impurities in CaSO₃ solids fed.

utilizations were significantly lower. (About 5% of the calcium sulfite fed remained in the effluent solids.) An important factor in the decreased H_2SO_4 utilization is the fact that as the incoming concentration of Na_2SO_4 is decreased while the concentration of Na_2SO_3 in the solution is held constant, proportionately more of the acid is utilized for the acidification of Na_2SO_3 than for the regeneration of Na_2SO_4 . Lowering the pH to 2.2 produced an even larger effective Na_2SO_4 removal but sulfuric acid utilization efficiency decreased even further.

Thus, as is not unexpected, removing a given amount of Na_2SO_4 per pass through the reactor will result in increasingly poorer utilization of sulfuric acid as the Na_2SO_4 level in the incoming stream decreases. As a consequence, for a fixed process oxidation rate, the Na_2SO_4 regeneration process will be increasingly inefficient as the required steady state Na_2SO_4 level in the process is lowered.

B. PILOT PLANT RESULTS

The evaluation of the sulfuric acid slipstream treatment process involved both open-loop reactor tests to confirm and extend the laboratory performance data, and the use of the sulfuric acid treatment system in conjunction with the complete dual alkali absorption/regeneration loop operating in a concentrated lime mode. In both sets of the tests, the sulfuric acid reactor system was operated in a similar manner. Filter cake slurry was fed continuously to the reactor together with 20 - 30 wt% sulfuric acid at a rate sufficient to maintain a prescribed reactor pH in the range of 2.3 to 3.3.

The filter cake slurry was made up as required on a batch basis. In the integrated system operation, a portion of the filter cake generated during the run was slurried with venturi recycle liquor. In the open-loop tests, the slurry was prepared using filter cake saved from prior dual alkali operations and either simulated venturi effluent liquor or simulated thickener underflow liquor.

The sulfuric acid reactor consisted of a baffled 30-gallon Pfaudler kettle vented to the atmosphere. The level in the reactor was adjusted to maintain approximately a 30-minute reactor holdup time. Effluent from the reactor was sent to a 6-inch solid bowl centrifuge for the separation of the solids. In the full system operation, the centrate was returned to the lime regeneration reactor.

A schematic of the overall system operation is provided in Figure IV-1.

1. Sulfuric Acid Reactor Performance

The results of the sulfuric acid reactor pilot plant tests confirmed the basic laboratory findings. The reaction proceeds readily to completion with formation of a gypsum product with good settling properties. Preliminary pilot tests showed that a 30-minute reactor holdup time is sufficient for essentially complete conversion; and the optimum pH appeared to be above 2.35.

Subsequent runs were then made over a range of pH from 2.6 to 3.3 in three different operating regimes: Na_2SO_4 limited, CaSO_3 limited, and approximately stoichiometric ratios of Na_2SO_4 to CaSO_3 . The results of these runs are summarized in Table IV-2 and Figure IV-2.

As would be anticipated, the runs with $\text{Na}_2\text{SO}_4/\text{CaSO}_3$ feed ratios nearest to stoichiometric (0.5) resulted in the highest reactor efficiencies. (Here, efficiency is defined as the decrease in mols of sodium sulfate divided by the mols of sulfuric acid used -- $[\text{mols } \text{Na}_2\text{SO}_4 \text{ in} - \text{mols total } \text{SO}_4 = \text{out}] / \text{mols } \text{H}_2\text{SO}_4 \text{ used}$.) The best efficiency attained in the pilot plant operations was 0.64 at a ($\text{Na}_2\text{SO}_4/\text{CaSO}_3$) feed ratio of 0.54. In cases where the feed ratio was greatly different from stoichiometric, the reactor efficiencies were found to decrease markedly. In the sulfate limited regime, acid is wasted dissolving CaSO_3 which is not used; and in the

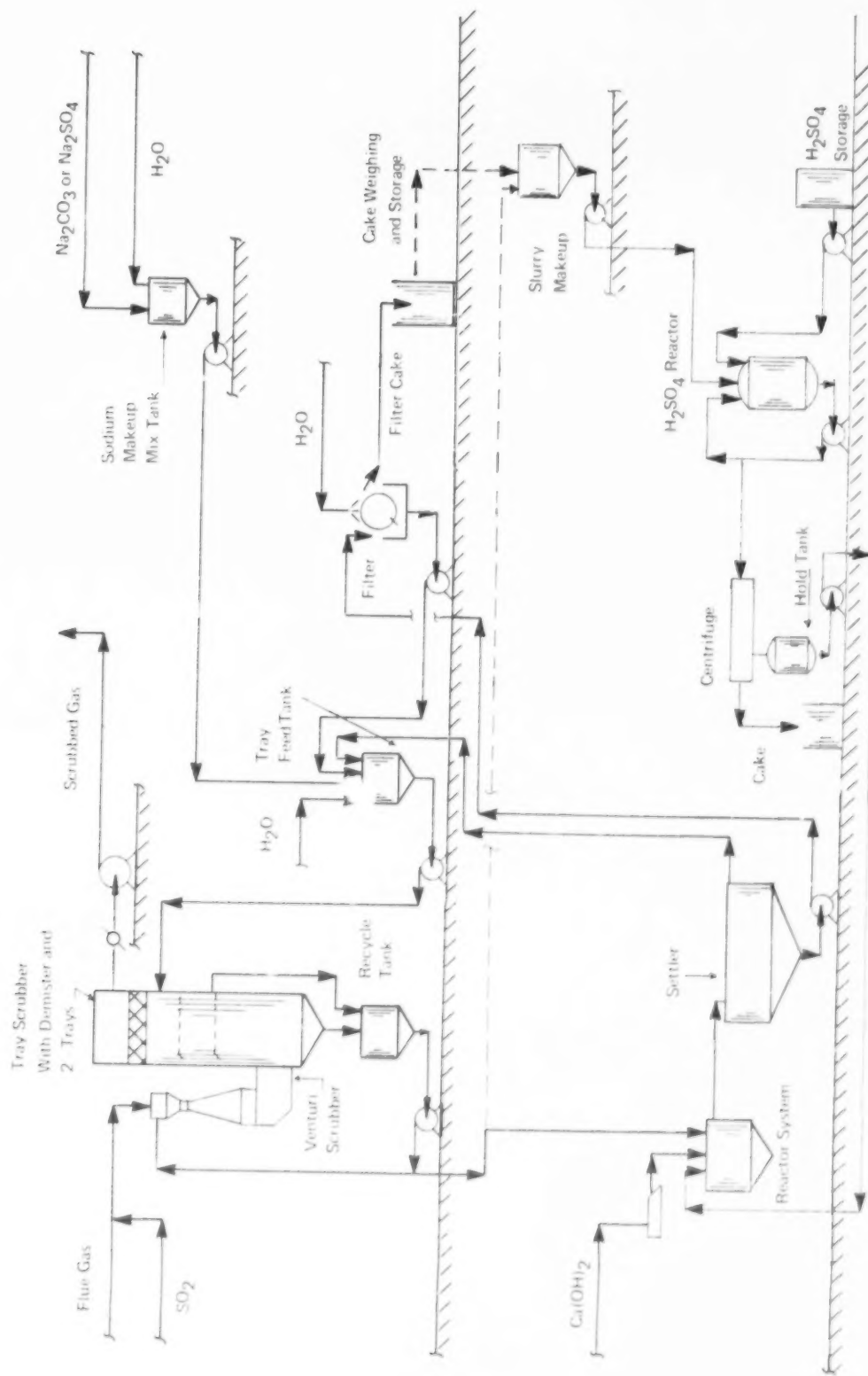


Figure IV - 1 Schematic Flow Diagram of Pilot Plant Operation For Sulfuric Acid Treatment Mode

Table IV-2 SUMMARY OF SULFURIC ACID SLIPSTREAM TREATMENT RESULTS

Sulfuric Acid Reactor Conditions: $\tau = 30$ min (CSTR) Residence Time
 $T = 80 - 110^\circ\text{F}$
 $\text{pH} = 2.6 - 3.3$

Feed Streams: Solids Source - Washed Filter Cake
 Liquor - Scrubber Bleed or Thickener Underflow
 Acid Strength = $2.5 - 3.0 \text{ M H}_2\text{SO}_4$

Operating Regime	Run No.	Feed Composition				Solids Composition, (wt%) CaSO_3 CaSO_4	mols Na_2SO_4 mol CaSO_3	Reactor Efficiency ^a	
		Wt % Solids	Liquid Composition $[\text{SO}_4]_L, \text{M}$	pH				Observed	Corrected ^b
Na_2SO_4 Limited	056	24.	0.63	11.5	74	10	0.32	0.22	0.33
	055	18.5	0.59	6.9	82	7	0.38	0.37	0.45
Stoichiometric	016-2	19.5	0.74	6.9	78	17	0.46	0.52	0.55
	016-1	19.	0.76	7.1	80	15	0.48	0.56	0.59
	016-3	19.5	0.83	6.9	76	18	0.54	0.64	0.68
	017-2	18.5	0.80	7.2	70	19	0.57	0.61	0.70
CaSO_3 Limited	050	13.	0.58	6.9	78	11	0.60	0.43	0.52
	017-1	18.5	0.90	7.1	70	19	0.67	0.39	0.44
	052	10.5	0.54	11.0	74	13	0.75	0.22	0.27
	051	11.	0.57	7.0	74	11	0.76	0.28	0.32
	053	26.5	1.07	11.5	36	52	0.95	0.10	0.13

^a $\frac{\text{mols Na}_2\text{SO}_4 \text{ in} - \text{mols total SO}_4 \text{ out}}{\text{H}_2\text{SO}_4} = \text{Efficiency}$

^b Efficiency corrected for NaOH , CaCO_3 and Na_2SO_3 alkalinity in feed slurry

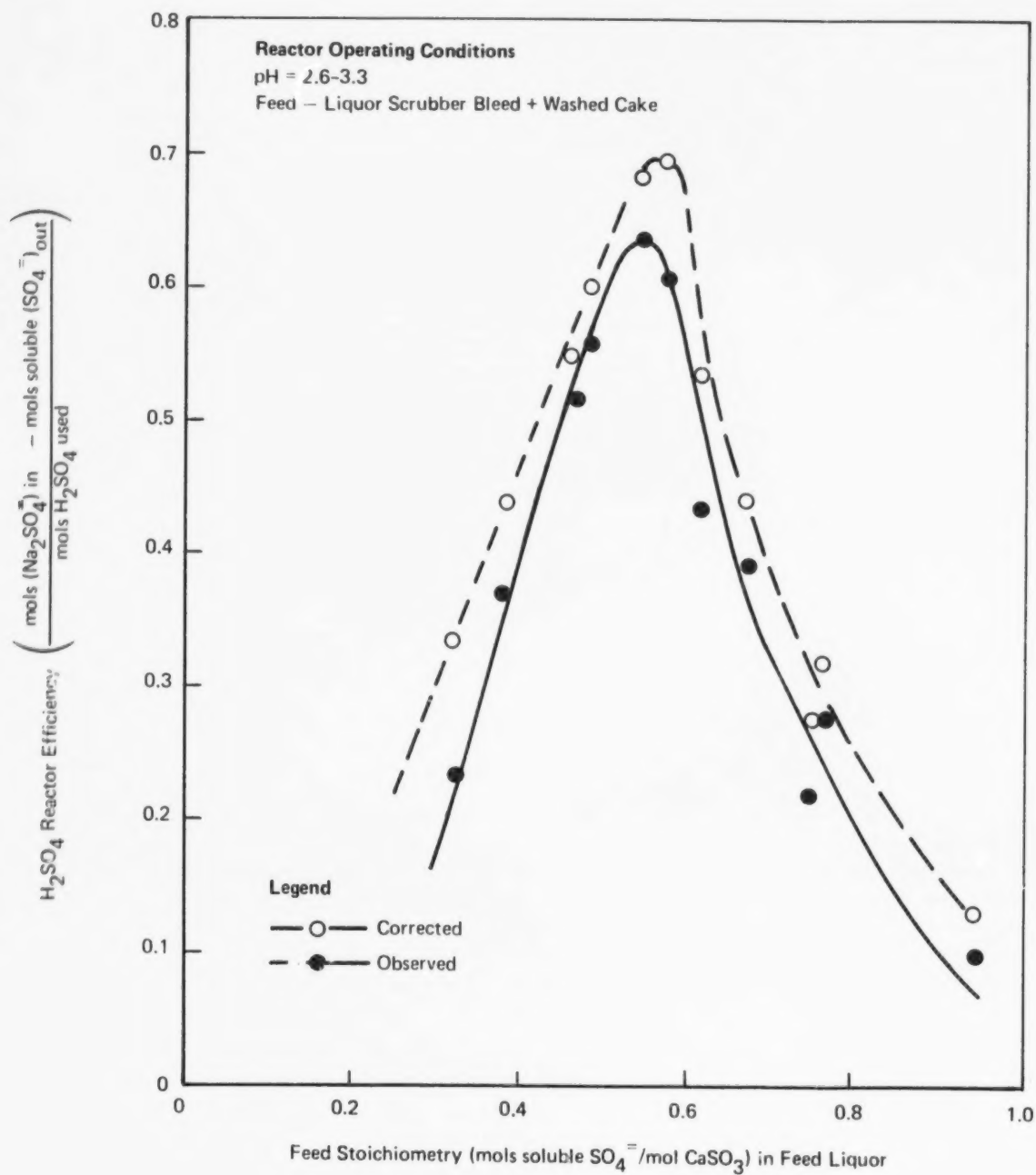


Figure IV-2 H_2SO_4 Reactor Efficiency versus Feed Stoichiometry

CaSO₃ limited regime, acid is wasted neutralizing and adjusting the pH of extra feed solution.

The pilot plant data plotted in Figure IV-2 cover a range of operating conditions including sulfate and sulfite concentration, reactor pH, ionic strength, and alkalinity content of the feed slurry. The chemistry of the system indicates that all of these factors will influence the efficiency of the reaction. Since all of these effects are confounded to some degree in the pilot plant results, the correlation of reactor efficiency with feed stoichiometry contains secondary factors other than analytical error that contribute to the data scatter. The efficiency/feed stoichiometry curve given in Figure IV-2 should be considered a generalized representation.

An attempt was made to extract the effects of feed alkalinity in order to clarify the observed performance. This involved adjusting the sulfuric acid used in each run to account for that acid used in neutralizing CaCO₃, NaOH and Na₂SO₃ in the feed. The "corrected" reactor efficiencies based upon these adjusted acid requirements are listed in Table IV-2 and are superimposed in Figure IV-2. As would be expected, the magnitude of the correction was generally greatest for the runs using simulated thickener underflow and for runs in the CaSO₃ limited regime where excess solution was fed.

The effects of sulfate concentration and reactor pH were specifically addressed in the laboratory program. This work showed that the optimum reactor operating pH is roughly 2.5 to 3.0, the range in which almost all pilot plant runs were made. Furthermore, as sulfate concentrations increase in the reactor feed slurry, reactor efficiency improves. This effect is not clearly demonstrated in the pilot plant data due to the great range of feed stoichiometry, which has a more pronounced influence on reactor performance.

Finally, the ionic strength of the reactor effluent slurry has a small effect on efficiency since it determines, along with temperature, the solubility product of gypsum. As ionic strength increases, the solubility product should also increase and, therefore, reactor efficiency should correspondingly decrease. Although the effect on reactor efficiency is too small to be observed, the increase in the gypsum solubility product with ionic strength was observed. Figure IV-3 shows the apparent solubility product curve for gypsum (calculated by molar concentrations) predicted by ADL along with solubility data generated in both the pilot plant and laboratory tests. For all but a few runs, the experimentally determined values of the apparent solubility product fall slightly below the predicted curve (predicted by the method of Kusik and Meissner), indicating that the solutions may not have been saturated with respect to CaSO₄·2H₂O. The validity of this predicted solubility product curve is demonstrated by the gypsum saturated data presented in Section VI.-A.

The pilot plant solubility data shown in Figure IV-3 correspond to soluble sulfate concentrations of 0.1M to 0.4M SO₄⁼ and calcium levels of 0.01M to 0.04M Ca⁺⁺ (400 to 1600 ppm). The lowest effluent sulfate levels, and therefore highest calcium levels, were achieved in cases where the reactor efficiency was highest.

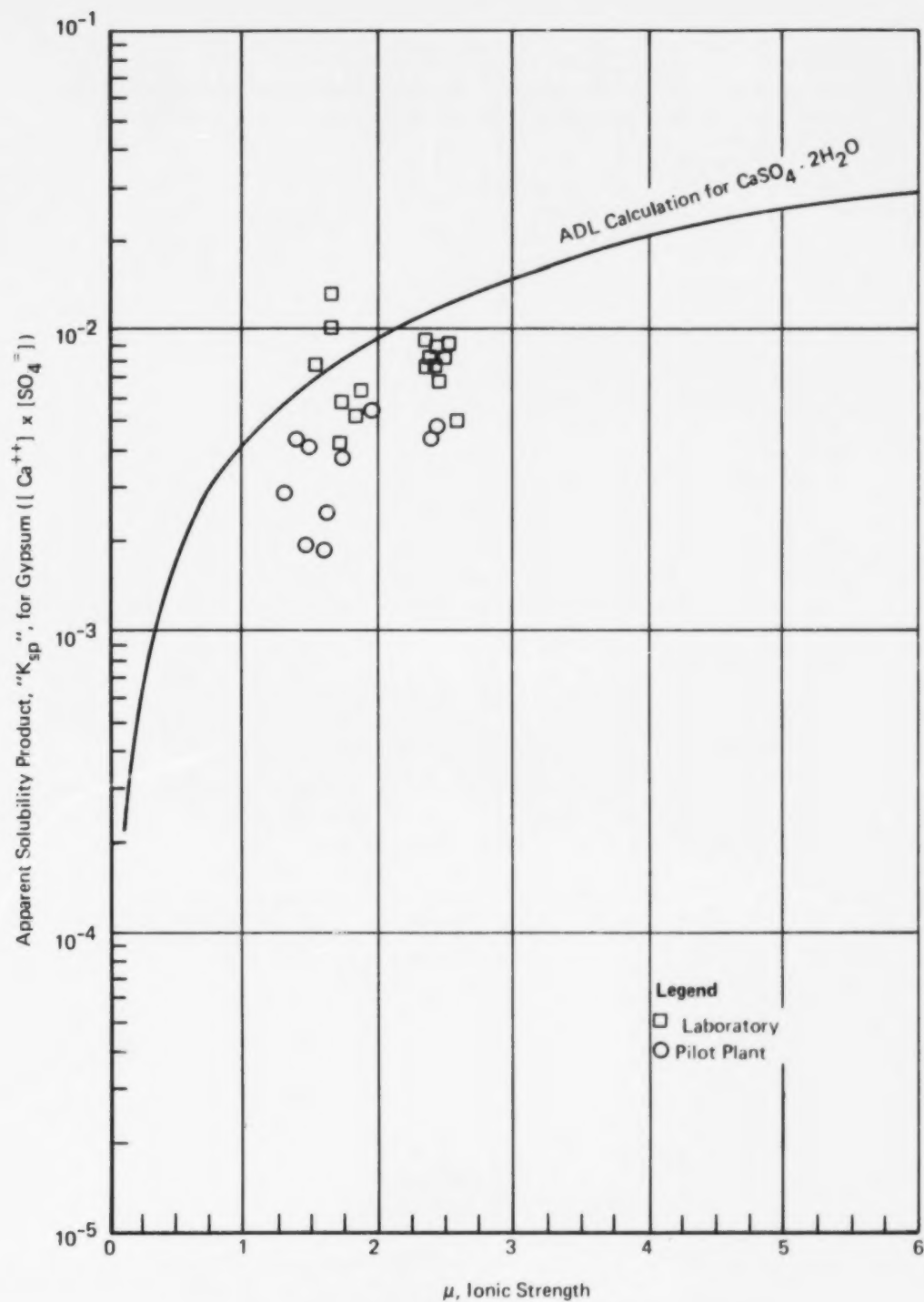


Figure IV-3 Extent of CaSO_4 Under-Saturation in H_2SO_4 Reactor Effluent – Pilot Plant and Laboratory Data

Although the highest efficiency achieved in the pilot plant tests was 0.64, it is reasonable to assume that refinements in the operating conditions and slight adjustments in the process configuration could raise efficiencies to above 0.7. For example, the reactor vent gas could be recovered and used for pre-acidifying the incoming filter cake slurry. Also, if Na_2SO_4 is used for the makeup sodium value, then this could be added either as a concentrate or solid directly to the sulfuric acid reactor feed slurry.

2. Integrated System Operation

Two closed-loop pilot plant tests were run for extended periods with the sulfuric acid reactor system operating on a slipstream from the dual alkali process. The overall system was operated in the concentrated active sodium mode using hydrated lime for absorbent regeneration. The first test, run 016, was made at an inlet SO_2 level of 2250 ppm. The second test, run 017, was conducted at an inlet SO_2 level of 650 ppm. The process conditions and overall results are summarized in Table IV-3.

In both runs the operation of the sulfuric acid treatment system was similar to the open-loop tests as previously described. Washed filter cake generated during the run was used as the CaSO_3 source and venturi recycle liquor provided Na_2SO_4 . The dual alkali process liquor was originally primed to a soluble sulfate level of about 0.8M to ensure that sufficient sulfate was present to produce high reactor efficiency. The rates of feed of cake slurry and sulfuric acid were initially set such that the total anticipated oxidation in the dual alkali system could be sustained by the combined sulfate precipitation in the absorbent regeneration reactor and sulfuric acid reactor and by the loss of soluble sulfate in the waste cake.

The performance of the sulfuric acid system in these runs has been discussed in connection with the sulfuric acid reactor performance in the previous section. Overall, the results were very good. Efficiencies ranged from 0.38 to 0.64 throughout both tests, and the average efficiency over each test was greater than 0.5. The gypsum produced had good dewatering properties. Centrifuge cake containing up to 70 wt% insoluble solids could be obtained by close control of the centrifuge feed rate.

The primary effect of the sulfuric acid slipstream treatment system on the overall process operation in these two runs was in the increased lime requirements (for ultimate neutralization of sulfuric acid added to the system). There were no adverse effects on calcium utilization, soluble calcium levels, or properties of solids generated in the absorbent regeneration reactor. In all of these respects, the performance of the system was consistent with that of a concentrated mode dual alkali process without sulfuric acid treatment.

The consumption of lime, however, rose significantly. In run 016 where the SO_2 removal equaled 2050 ppm, sulfite oxidation throughout the system amounted to about 570 ppm, or 28% of the SO_2 removal. Approximately half of this oxidation rate was sustained by sulfate precipitation in the lime reactor

Table IV-3 SUMMARY OF CLOSED-LOOP RUNS

Sulfuric Acid Treatment Mode

Run No.	016	017
<u>Dual Alkali Process Operation</u>		
Inlet SO ₂ Level, ppm	2250	650
SO ₂ Removed, ppm	2050	530
Scrubber Bleed: pH	6.1	6.1
[TOS], M	0.35	0.5
[SO ₄ ⁼], M	0.7	0.8
Overall System Oxidation		
Equivalent ppm SO ₂ (avg.)	570	400
% ΔSO ₂ (avg.)	28%	70%
Sulfate Precipitation in Lime Reactor:		
Equivalent ppm SO ₂ (avg.)	250	100
% ΔSO ₂ (avg.)	12%	18%
Calcium Feed:		
mols Ca(OH) ₂ /mol ΔSO ₂	1.33	1.83
mols Ca(OH) ₂ /mol sulfur input	1.06	1.01
<u>H₂SO₄ Treatment System Operation</u>		
H ₂ SO ₄ Feed, (mols H ₂ SO ₄ /mol ΔSO ₂)	0.26	0.63
Reactor Efficiency	0.52-0.64	0.38-0.61
Sulfate Precipitation:		
Equivalent ppm SO ₂	280	180-300
% of ΔSO ₂	14%	34-57%
Centrifuge Cake Composition:		
Insoluble Solids (% total wt)	60-70	53
Soluble Solids (% dry cake)	4-7	15

and by soluble sulfate losses in the waste cakes. The remainder, about 14% of the SO_2 absorbed, was removed in the sulfuric acid reactor. Since the reactor efficiency averaged 0.57 (efficiencies ranged from 0.52 to 0.64), the sulfuric acid requirement amounted to about 26% of the SO_2 absorbed. This acid feed was directly reflected in the required increase of 26% in the lime feed ($1.33/1.06=1.26$). At the lower SO_2 inlet level of run 017, the use of sulfuric acid increased markedly due primarily to oxidation in the dual alkali system being a higher percentage of the SO_2 absorbed. In addition, Na_2SO_4 was used to make up lost sodium value. The combined oxidation and sulfate addition resulted in an increase in the lime feed of 80 - 85% over that required for SO_2 removal alone; and a total sulfuric acid requirement of 63% of the SO_2 removed.

It is clear from the pilot plant results that the sulfuric acid treatment scheme is a technically feasible and reliable approach to the removal of soluble sulfate from dual alkali systems. Because the use of this sulfuric acid treatment scheme may be costly when applied to systems with high oxidation rates ($>40\%$ of ΔSO_2) due to the sulfuric acid and extra lime requirements, it may be more appropriate for systems with intermediate levels of oxidation where the rate of sulfate formation cannot be easily handled in a simple concentrated active sodium mode.

Composite Diagram of Projected Pilot Plant Operations Figure IV-4 summarizes the operational characteristics of the various dual alkali process sections for the concentrated lime mode operation with sulfuric acid slip-stream treatment for a high SO_2 inlet condition. This composite is based upon pilot plant operations and is analogous to the block diagrams provided in Section III. C. for the simple concentrated lime mode. Figure IV-4 indicates the various inputs and outputs expressed in terms of SO_2 absorbed for a dual alkali system with an overall oxidation rate equivalent to 30% of the SO_2 absorbed. For each of the four process sections, the observed rates of sulfite oxidation and/or active sodium regeneration are also expressed as a function of SO_2 absorption.

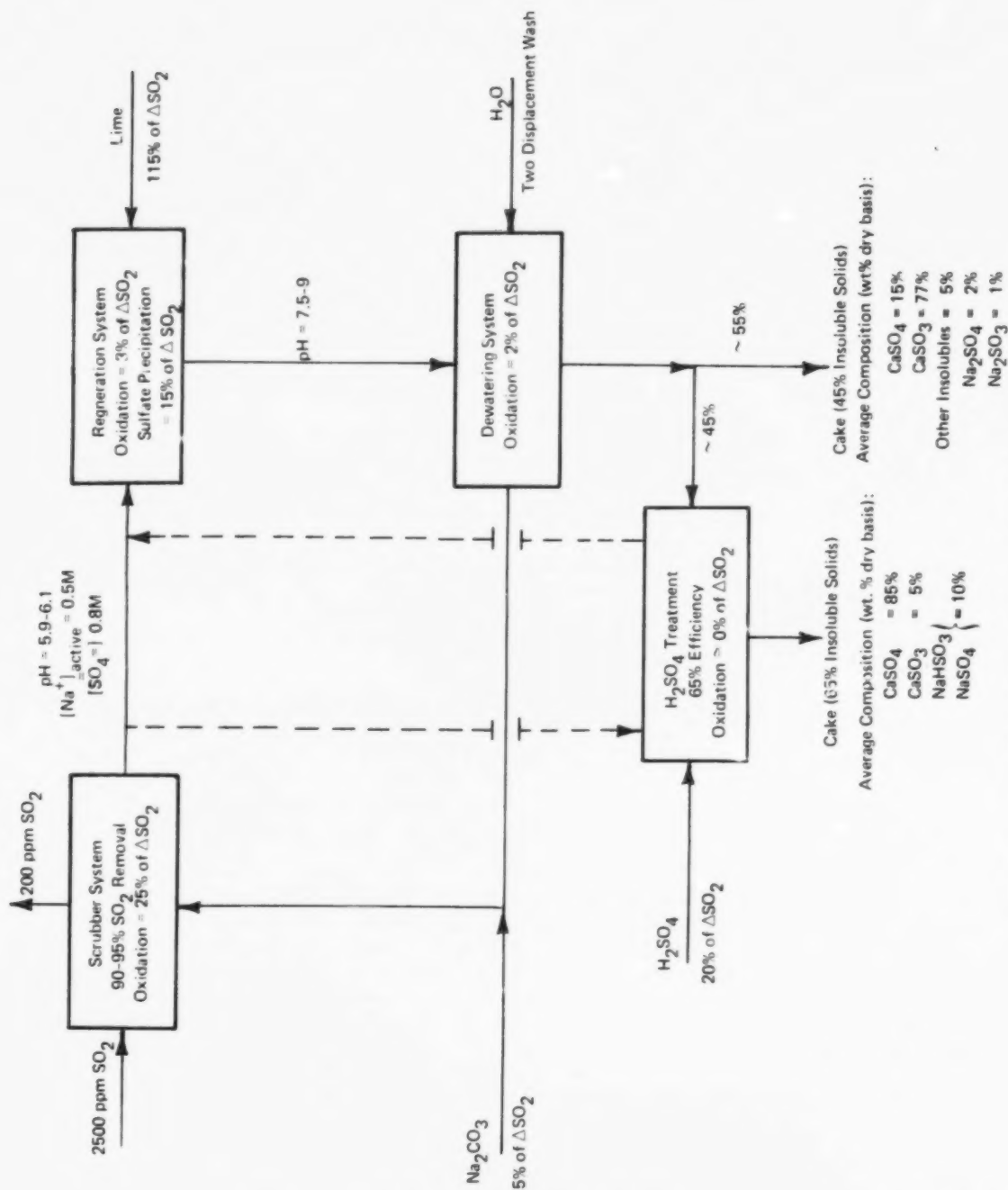


Figure IV-4 Composite Pilot Plant Operation - H₂SO₄ Treatment Mode

C. SULFURIC ACID REACTOR MODEL

The model of the sulfuric acid reactor system was developed to simulate the effects of various dual alkali process conditions on the sulfuric acid system efficiency. The principal purpose of the model was to provide a method for estimating reasonable sulfuric acid reactor operating conditions and for determining the sulfuric acid requirements in different dual alkali system applications. As developed, the model applies to the operation of the reactor in the CaSO_3 limited regime. It is based upon the reaction equations discussed in Section IV.-A. and assumes that chemical equilibrium is achieved in the system. The model determines the amount of sulfuric acid required to neutralize all alkalinity values in the slurry feed and adjust the pH into the 2.3 to 3.3 range. The levels of sulfate and calcium in solution in the reactor effluent are determined from an estimate of the " K_{sp} " for CaSO_4 (" K_{sp} " = $[\text{Ca}^{++}] \times [\text{SO}_4^{--}]$ with concentrations in mols/liter). The value of " K_{sp} " used is the average of those values calculated from the laboratory data for the range of ionic strengths expected in the pilot plant operations. Since " K_{sp} " is an input to the model, this can be adjusted for any desired ionic strength.

In order to adjust for pH, it has simply been assumed that enough sulfuric acid is added to convert about 20% of the system TOS to H_2SO_3 . This is a rough estimate from the laboratory results. It would, of course, be possible to try to determine the $\text{H}_2\text{SO}_3/\text{HSO}_3^-$ ratio from pK for H_2SO_3 . However, very little data exist which will allow for reasonable estimates of the reference activity coefficients for the H_2SO_3 and HSO_3^- species at ionic strengths greater than 1.0. Furthermore, initial tests with the model indicate that calculated reactor efficiencies were relatively insensitive to rather large changes in this pH adjustment factor.

No consideration has been specifically given in the model to the formation of HSO_4^- . At the pH's involved, the amount of HSO_4^- should be less than a few percent of the total S(VI) species.

Model Application The model has been used primarily to design pilot plant experiments and to determine the sulfuric acid requirement for various dual alkali system oxidation rates.

Table IV-4 compares the experimental reactor efficiencies from the pilot plant operations with those predicted by the model. The model tracks the experimental runs fairly closely and, in most cases, the predicted value of efficiency is within about 15% of that observed. This agreement is good, particularly in light of the sensitivity of the model to the solids level in the feed slurry. A change of 10% in the solids level (from 20% to 22% solids) changes the efficiency estimate by about 30% (from say 40% to 52%). Since the solids level is an experimentally determined value that can vary slightly during a run, this sensitivity is significant.

The effects of other operating conditions on the reactor efficiency are smaller. Small changes in $[\text{OH}^-]$, $[\text{TOS}]$ and CaCO_3 in the cake produce only minimal changes in efficiency. The effect of changes in sulfate concentration

Table IV-4 MODEL SIMULATIONS OF PILOT PLANT OPERATIONS

Feed Ratio ($\frac{\text{mols Na}_2\text{SO}_4}{\text{mol CaSO}_3}$)	Experimental Reactor Efficiency	Efficiency Predicted By Model
0.54	0.64	0.72
0.57	0.61	0.56
0.60	0.43	0.45
0.67	0.39	0.43
0.75	0.22	0.13
0.76	0.28	0.19
0.95	0.10	-0.12

also tend to be small, although to some extent this is a function of the actual level of sulfate. As sulfate concentrations in the feed decrease toward the equilibrium effluent sulfate concentrations, changes in sulfate concentration become more important.

Neither the pH adjustment factor nor the value of " K_{sp} " has been found to have a strong influence on the model performance. A 100% increase in " K_{sp} " from 8×10^{-3} to 16×10^{-3} generally decreases the predicted reactor efficiency by less than 10%. Similarly a 50% increase in the pH adjustment factor, from 0.2 to 0.3, usually decreases the efficiency prediction by only about 5%.

D. APPLICABILITY OF THE SULFURIC ACID TREATMENT SYSTEM

In using a sulfuric acid slipstream treatment system, the amount of sulfuric acid required is important because it directly affects the overall lime requirement. As the sulfuric acid rate increases, the lime rate must increase accordingly to remove the additional sulfur value added to the system. For example, in order to precipitate sulfate at a rate sufficient to keep up with an oxidation rate equivalent to 15% of the SO_2 absorbed, the lime feed requirement will increase by 25% for a 60% reactor efficiency.

The implications of calcium utilization in the absorbent regeneration reactor on the sulfuric acid efficiency are also important, particularly if limestone is used for regeneration with lower expected calcium utilization. The influence of calcium utilization on reactor efficiency is illustrated in Figure IV-5. In order to ensure 30% reactor efficiency under reasonable system operating conditions, calcium utilization must exceed 75%; and in order to achieve 50% reactor efficiency the utilization must exceed 90%.

The consequences of using the sulfuric acid slipstream treatment approach for sulfate regeneration should, therefore, be carefully evaluated in terms of the overall process operation. In many cases where oxidation rates are high enough that they cannot be easily handled by a normal concentrated mode operation, other dual alkali approaches might be more promising than a sulfuric acid treatment scheme.

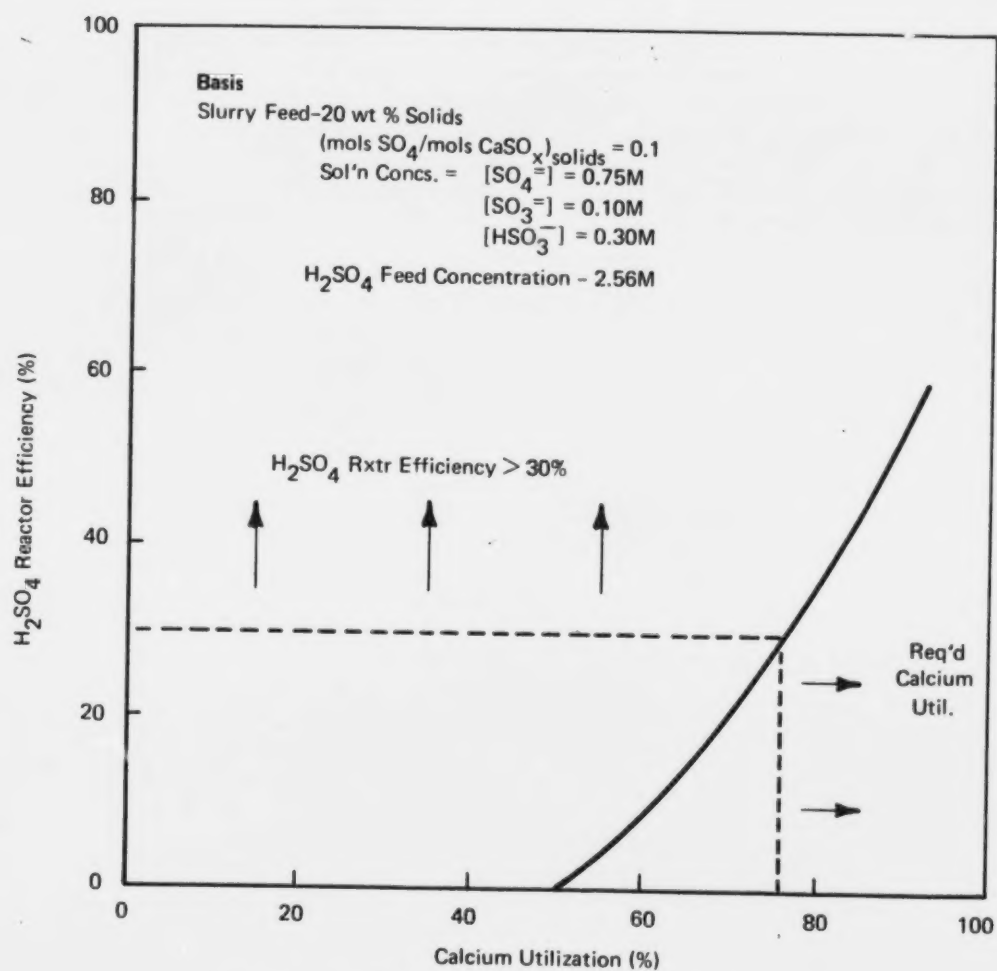


Figure IV-5 H_2SO_4 Reactor Efficiency Versus Calcium Utilization

V. LIMESTONE REGENERATION -- CONCENTRATED MODE

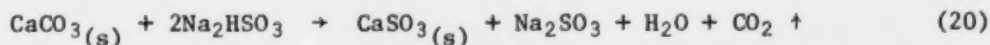
A. INITIAL LABORATORY STUDIES

Since the incremental cost of lime over the equivalent amount of limestone is tied quite closely to the increasing cost of the energy required to calcine the limestone to lime, a considerable operating cost saving would be realized if limestone could be used directly in a dual alkali process in lieu of lime. Limestone has been used in FGD systems which employ direct slurry scrubbing, and the chemical equilibria involved have been characterized extensively. By analogy, limestone should be applicable to dual alkali regeneration. However, because there are significant differences between reaction conditions in direct slurry scrubbing and dual alkali regeneration -- one of the most important is the significantly higher ionic strengths found in concentrated dual alkali solutions -- a program of laboratory experiments was undertaken to study the reaction between limestone and concentrated sodium sulfite/bisulfite solutions. In a parallel effort, the use of limestone to regenerate dilute sodium sulfite/bisulfite solutions was studied by personnel at the Research Triangle Park laboratories of the EPA. The results of those studies are discussed in Section VI.

Limestone is being used in two commercial dual alkali systems in Japan. However, both systems utilize special treatment steps for sulfite precipitation; both ultimately produce gypsum and are relatively complex processes. They are described in more detail in a paper by Kaplan.⁷

The ADL laboratory program, still in progress, has consisted of an initial series of batch experiments in which the effects of the type of limestone, limestone feed stoichiometry, reaction temperature, and sodium sulfate level on the kinetics and equilibria of the reaction were studied. In subsequent continuous reactor experiments, limestone utilization and the physical properties of the product solids were studied as a function of a number of experimental variables.

As pointed out in the introduction, limestone will not react with sodium sulfite solutions because limestone is less soluble than the calcium sulfite which would be produced. But the bisulfite ion, HSO_3^- , is sufficiently acidic to react with limestone. This neutralization reaction, which results in the ultimate precipitation of calcium sulfite, can be written simplistically as follows:



In fact, the bisulfite initially present is never completely neutralized because at pH's where the reaction will proceed (up to about pH = 7), the species H_2CO_3 , HCO_3^- , HSO_3^- , and $\text{SO}_3^{=}$ all exist in significant quantities in equilibrium with one another. Nevertheless, based on the above equation one can express the amount of limestone fed as the percent of the stoichiometric amount which would be required to neutralize the HSO_3^- initially present.

$$\% \text{ Stoich} = \left[\frac{\text{mols CaCO}_3}{0.5 \times \text{mols HSO}_3^-} \right] \times 100\%$$

Thus, a feed stoichiometry of 100% is obtained when the mole ratio of CaCO_3 to HSO_3^- fed is 1:2. This definition of feed stoichiometry will be used throughout the discussion which follows.

Regardless of the nature of the species in solution, it is the removal of soluble sulfur-containing ions -- both sulfur (IV), sulfite; and sulfur (VI), sulfate -- from solution as insoluble calcium salts by which the effectiveness of limestone utilization is realistically evaluated. In CSTR experiments, utilization was based on analyses of the reactor product solids and calculated as follows:

$$\% \text{ Utilization} = \left[\frac{\text{mols CaSO}_4 \text{ in solids}}{\text{total mols Ca in solids}} \right] \times 100\%$$

Since the initial batch experiments were comparative in nature, for the sake of expediency the precipitation of CaSO_4 was ignored and limestone utilization was computed on the basis of the decrease in the concentration of soluble sulfur (IV) -- Total Oxidizable Sulfur [TOS] -- in solution as the reaction proceeded. Thus, if limestone is fed at 100% of stoichiometry and it is completely utilized, one would expect [TOS] to decrease by an amount equal to one half the concentration of bisulfite ion, $[\text{HSO}_3^-]$, initially present. This relationship can be expressed as follows:

$$\% \text{ Utilization} = \left[\frac{[\text{TOS}]_{\text{in}} - [\text{TOS}]_{\text{out}}}{0.5 \times [\text{HSO}_3^-]_{\text{in}} \times (\% \text{ Stoich}/100)} \right] \times 100\%$$

Because the formation of CaSO_4 is ignored in this calculation, it leads to conservative estimates of actual utilization.

1. Batch Studies Comparing Limestone Reactivity

The initial experiments in the laboratory program compared the relative reactivities of several types of limestone toward concentrated, acidic sodium sulfite/bisulfite solutions. In previous studies related to limestone slurry scrubbing, Drehmel⁸ found significant differences in reactivity which were attributed to differences in both the chemical and physical properties of the wide variety of limestones tested.

Materials from three sources were available in quantity for testing and subsequent use in this laboratory program. By chemical analysis all were more than 95% CaCO_3 . The materials included:

- Reagent grade CaCO_3 (precipitated chalk); Fisher Scientific Co.
- A natural calcite obtained from Pfizer, MPM Div., Clifton, New Jersey. This material (Marblewhite 200) was an industrial grade, low magnesium, limestone produced at Adams, Mass.
- Fredonia limestone ground by and obtained from the EPA/TVA Shawnee Test Facility.

Based on the sieve analyses shown below, the Marblewhite 200 and the Fredonia limestones appeared to be similar in their particle size distributions while the reagent CaCO_3 was uniformly finer.

Mesh	Limestone Type		
	Reagent	Marblewhite 200	Fredonia
- 300	100%	87.5%	85.7%
-200 + 300	-	11 %	8.9%
-100 + 200	-	1.5%	3.7%
- 60 + 100	-	0.2%	0.4%

Examination under the scanning electron microscope (SEM) revealed more subtle differences which were not evident from the sieve analysis. The reagent grade material was composed of quite uniformly sized cubes about 8 microns (μ) on a side. The Marblewhite particles also appeared crystalline but were very irregular and jagged in shape with a longest dimension ranging from 2 μ to greater than 50 μ ; an average length was estimated to be about 20 μ . The Fredonia material, which was similar to Marblewhite by sieve analysis, appeared very different under SEM examination. It appeared amorphous; the bulk of the particles were rough, irregular spheres about 2 - 8 μ in diameter. Particles larger than 10 μ were present but they appeared to be agglomerates of many smaller particles.

The results of batch reactions of the three limestones with similar concentration sodium sulfite/bisulfite sulfate solutions at 50°C are shown in Figure V-1. The progress of the reaction as a function of time was followed by measuring the decrease in TOS concentration in solution. The substantially higher reactivity of Fredonia limestone (as compared to reagent CaCO_3) is immediately evident from the more rapid TOS reduction obtained with the Fredonia material. The solution with which the Marblewhite was reacted had a somewhat higher initial TOS level than the other two solutions.

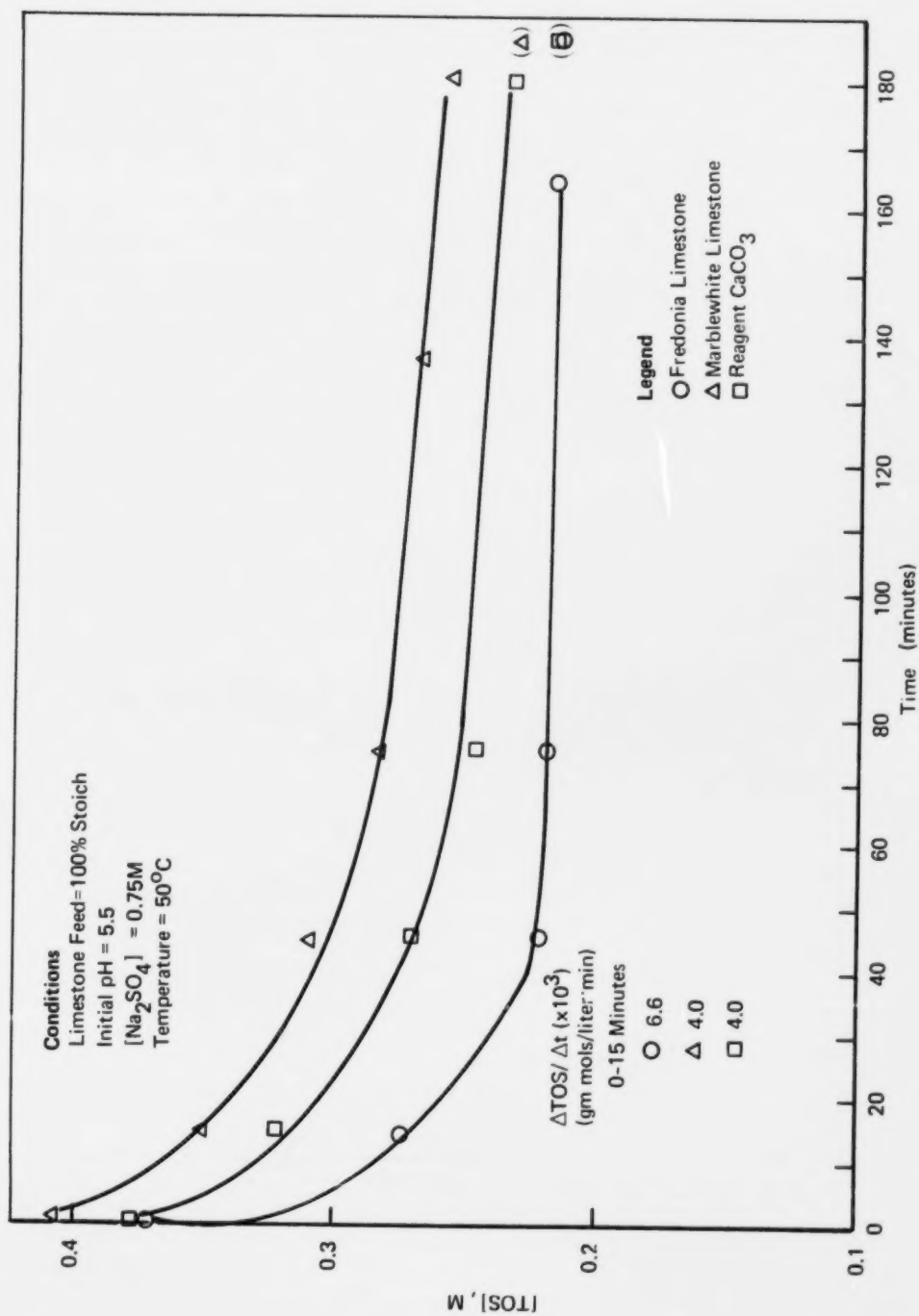


Figure V-1 Comparison of Limestone Reactivities in Batch Reactions

To permit a better comparison of reactivity, initial reaction rates were computed and the values included in Figure V-1. The overall reaction rates shown are the change in TOS concentration per minute for the reaction time period 0 - 15 minutes. The computed overall reaction rates for Marblewhite limestone and reagent CaCO_3 were essentially the same while the rate observed for Fredonia limestone was about 50% faster than either of the other two.

With the Fredonia limestone, TOS changed very little after the first 45 minutes of reaction. With the other two materials, a noticeable reduction in TOS, indicative of continuing slow reaction, was observed throughout the remainder of the experiments which were terminated after three hours.

Plotted at the extreme right-hand side in parentheses in Figure V-1 are the TOS levels which would have been obtained if the limestone which was fed had been completely utilized to precipitate calcium sulfite. The final TOS level observed for Fredonia limestone was much closer to the theoretical level than it was for the other two materials. Utilizations, based on TOS reduction, in solution, even after three hours of reaction, were markedly poorer for Marblewhite limestone and reagent CaCO_3 (81 and 85%, respectively) than for the Fredonia material (96%). Utilization of the Fredonia material after only 45 minutes of reaction was about 92%, whereas utilizations of Marblewhite and reagent CaCO_3 after 45 minutes were only 54 and 64%, respectively. If precipitation of CaSO_4 had been taken into account, all utilizations would have been higher; in the case of Fredonia, utilization would have been about complete.

Because of the chemical similarity of all three materials, the observed difference in reactivity between Fredonia and the other two materials probably resulted from the fact that the Fredonia limestone consisted of amorphous particles with a higher accessible surface area than the crystalline particles characteristic of the other two.

Because of its significantly higher reactivity in batch tests, the Fredonia limestone was used almost exclusively in the remainder of the batch and CSTR experiments that were performed. We do realize, however, that further consideration must be given to the effect of limestone reactivity before a complete assessment of its potential for dual alkali regeneration in a wide variety of situations can be made. The batch reaction method employed seems to be a relatively simple but realistic means to study reactivities. Additional tests of the reactivity of materials from a variety of locations are planned as part of our continuing ADL laboratory program.

2. Effects of Feed Stoichiometry, Sulfate Concentration, and Temperature on Reaction Rate

The effect of changing the Fredonia limestone feed stoichiometry on the rate of the regeneration reaction is shown in Figure V-2. When the amount of limestone fed was reduced from 100% to 50% of the stoichiometric amount

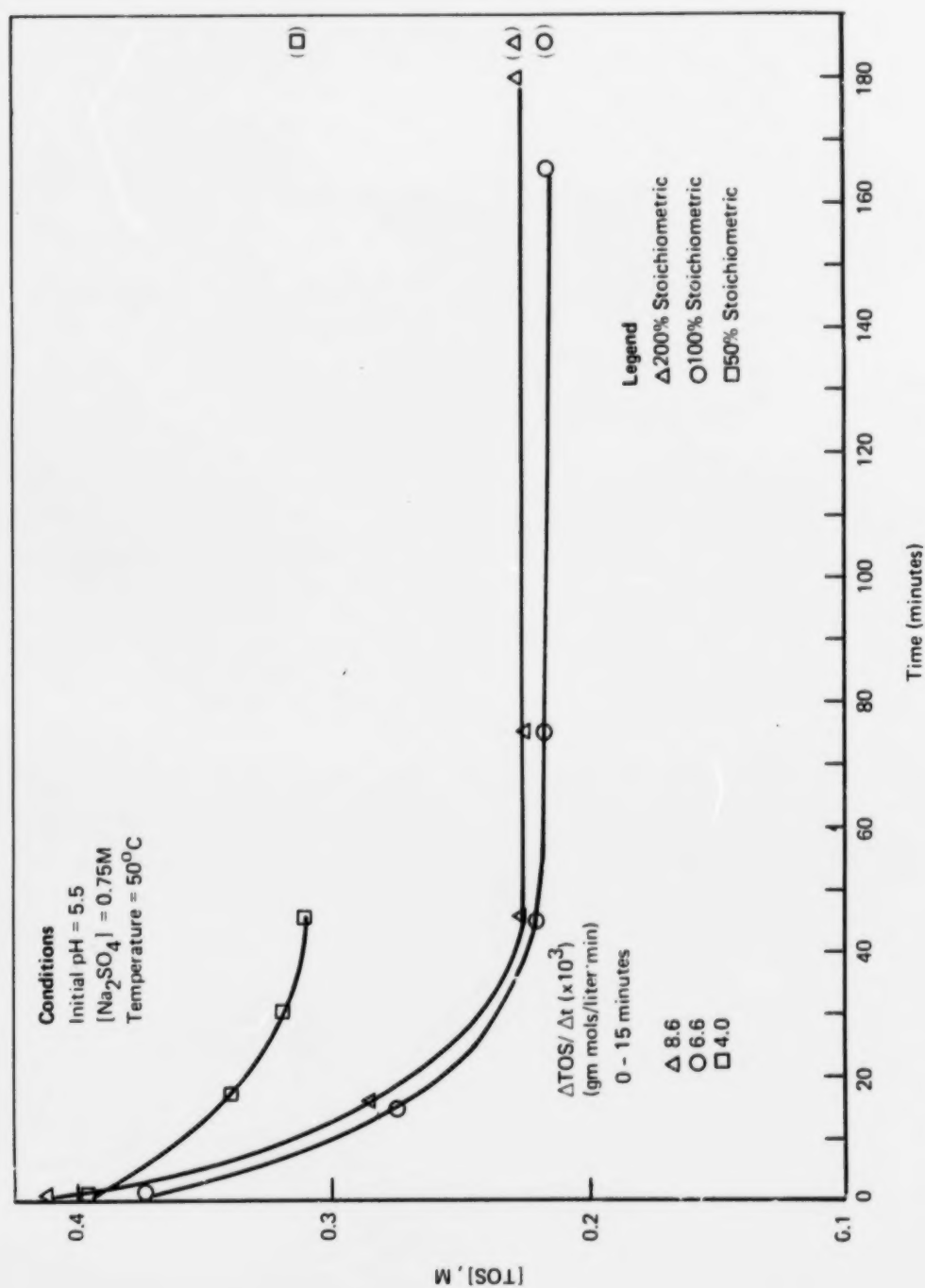


Figure V-2 Effect of Limestone Feed Stoichiometry on Regeneration Rate With Fredonia Limestone

required to neutralize the HSO_3^- initially present, the reaction rate during the first 15 minutes of reaction also decreased significantly, although the percent decrease in reaction rate was not as great as the reduction in the feed stoichiometry. The theoretical minimum TOS level after 45 minutes of reaction was more closely approached at the lower (50%) stoichiometry; and the resulting limestone utilization (based on TOS reduction only) was about 97% as compared to 92% when the feed was 100% of stoichiometric. When the limestone feed stoichiometry was doubled from 100% to 200%, the initial reaction rate increased about 30% and the TOS concentration after 45 minutes had fallen essentially to the minimum theoretical level. Within experimental error, the concentration of TOS remained constant during the remainder of the experiment. However, none of the limestone added in excess of 100% stoichiometric reacted -- the overall utilization was only about 50%.

The dependence of the reaction rate on limestone feed stoichiometry observed in these experiments indicates that limestone dissolution is one of the rate limiting steps in the regeneration reaction with limestone.

The effect on reaction rate of the concentration of Na_2SO_4 in solution is shown in Figure V-3. It is immediately obvious that at 1.25M Na_2SO_4 , the reaction was significantly slower than at the two lower Na_2SO_4 levels. Examination of the computed reaction rates for the initial 15 minutes of reaction reveals that initial rates were significantly different for each of the Na_2SO_4 levels studied. Each increase in the concentration of Na_2SO_4 was accompanied by a corresponding decrease in initial reaction rate. However, the reaction rate for the first 45 minutes was essentially the same for Na_2SO_4 levels of 0.25M and 0.75M, but the corresponding rate at 1.25M Na_2SO_4 was still significantly lower.

All of the preceding experiments were conducted at a temperature of 50°C which is representative of the temperature at which a reactor would operate if scrubber effluent were neither heated nor cooled prior to regeneration. The significant decrease in reaction rate at a lower temperature, 38°C, is shown in Figure V-4. The 12°C reduction in reactor temperature produced essentially the same reduction in the calculated reaction rate as did increasing the concentration of Na_2SO_4 from 0.75M to 1.25M in the preceding experiment.

3. Continuous Reactor Studies of Regeneration with Limestone

The results of the batch experiments indicated that Fredonia limestone could be used effectively to regenerate concentrated sodium bisulfite/sulfite solutions. The regeneration with limestone was studied further in the laboratory CSTR (continuous, stirred tank reactor) to observe the physical properties of the solids produced, to ascertain whether or not sulfate would also be precipitated during the course of the reaction, and to obtain a more realistic estimate of achievable limestone utilization.

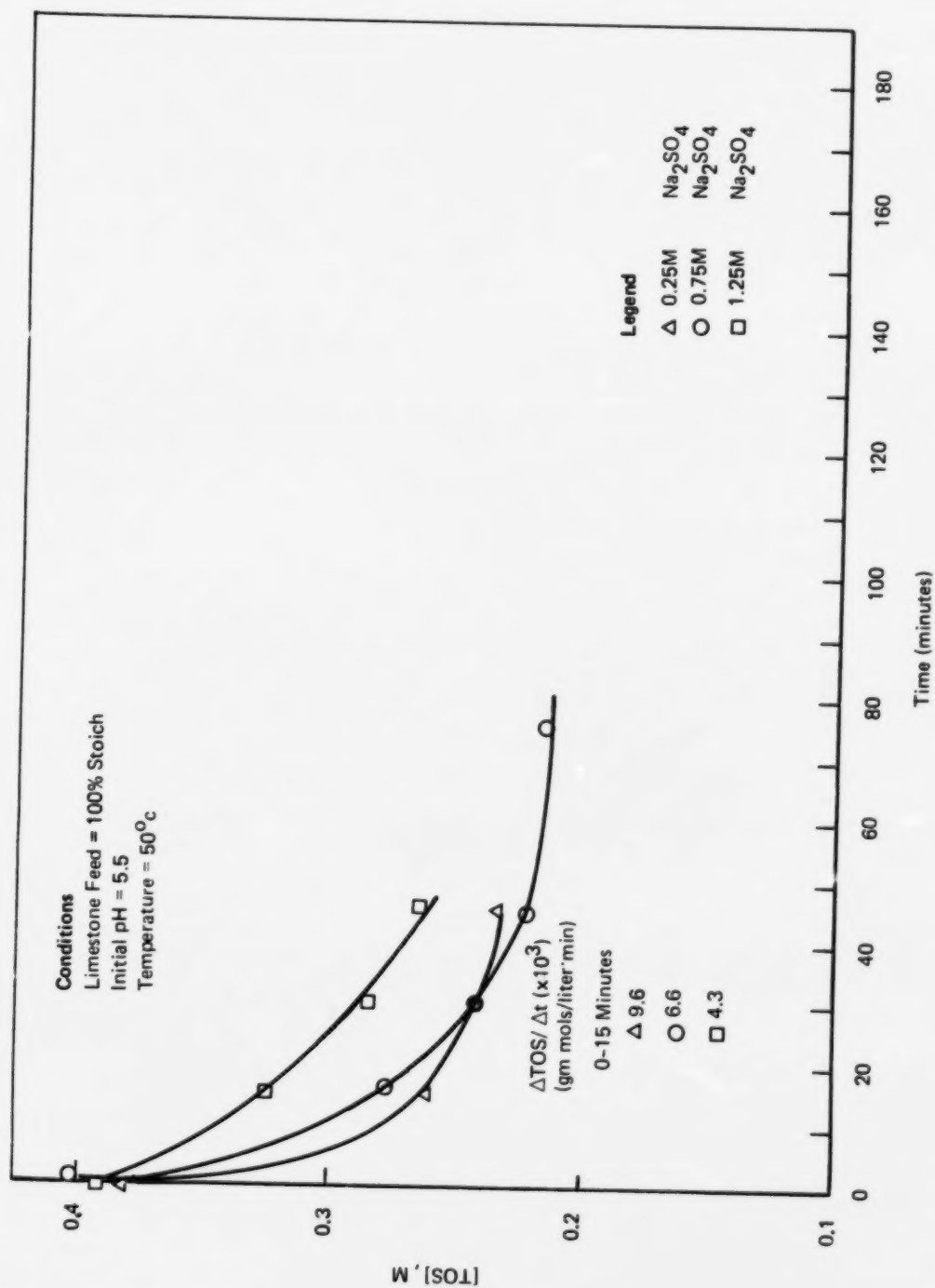


Figure V-3 Effect of Na₂SO₄ Concentration on Regeneration Rate With Fredonia Limestone

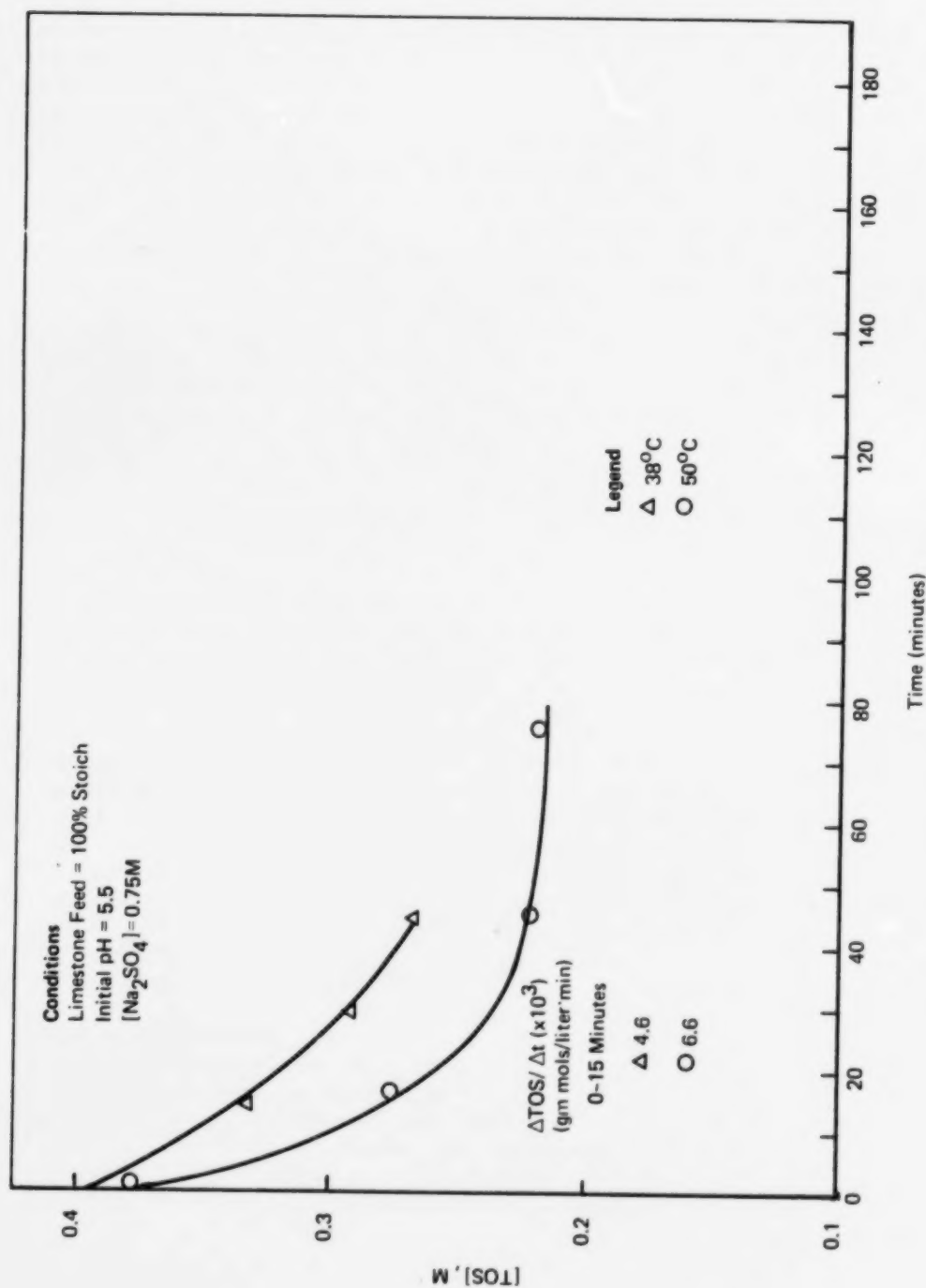


Figure V-4 Effect of Temperature on Regeneration Rate With Fredonia Limestone

The results of the batch experiments indicated that with Fredonia limestone, achieving nearly complete utilization at a temperature of 50°C and with Na_2SO_4 concentrations of 0.75M or less, required about 45 minutes of batch reaction time. In a CSTR that length of reaction would translate to a residence time of roughly 2 - 2.5 hours. However, the CSTR system available when these studies needed to be performed had been specifically designed to operate at shorter residence times (5-50 minutes) for studying regeneration with lime, and would not operate reliably at longer residence times. For the sake of expediency, the initial set of CSTR studies with limestone was performed at the 50-minute maximum residence time obtainable; a new CSTR is being constructed for additional studies at longer residence times.

The observed utilizations of limestone (based on CaSO_4 /total Ca in the product solids) in the 50-minute CSTR as a function of limestone feed stoichiometry are shown in Figure V-5. Included for reference are the utilizations achieved after 45 minutes of batch reaction based on TOS removal only; these are known to be somewhat low because CaSO_4 formation is not included. Even though the CSTR residence time was short, utilizations of Fredonia limestone greater than 80% were achieved with a feed stoichiometry of about 60%. When the stoichiometric amount of limestone fed was increased, utilization decreased significantly.

Utilization of the less reactive Marblewhite limestone was also studied in the 50-minute CSTR at a feed stoichiometry of 100%. Utilization of this material was only about 55%; utilization of Fredonia under similar conditions was about 70%.

Included in Figure V-5 are the results of a number of experiments in which the CSTR residence time was maintained at 50 minutes but a portion of the reactor effluent slurry was collected, filtered, and the solids recycled to the reactor. In once-through CSTR operation, the weight percent suspended solids in the reactor effluent slurry was typically about 2%. In several experiments with Fredonia limestone, the suspended solids level in the reactor was raised to about 5% by recycling solids; and as shown in Figure V-5, utilization of the limestone increased significantly over that observed for once-through operation. Operation with solids recycle also appeared to increase the utilization of Marblewhite limestone. The two experiments with Marblewhite were conducted at different feed stoichiometries, but when effluent solids were recycled, Marblewhite utilization was comparable to that obtained during once-through operation with Fredonia; in once-through operation, Marblewhite utilization had been significantly lower than that observed for Fredonia. Recycling solids effectively increased the amount of time available for the limestone to react by a factor of 2.5 - 3. A similar increase in reactor residence time should result in a similar improvement in utilization for once-through operation.

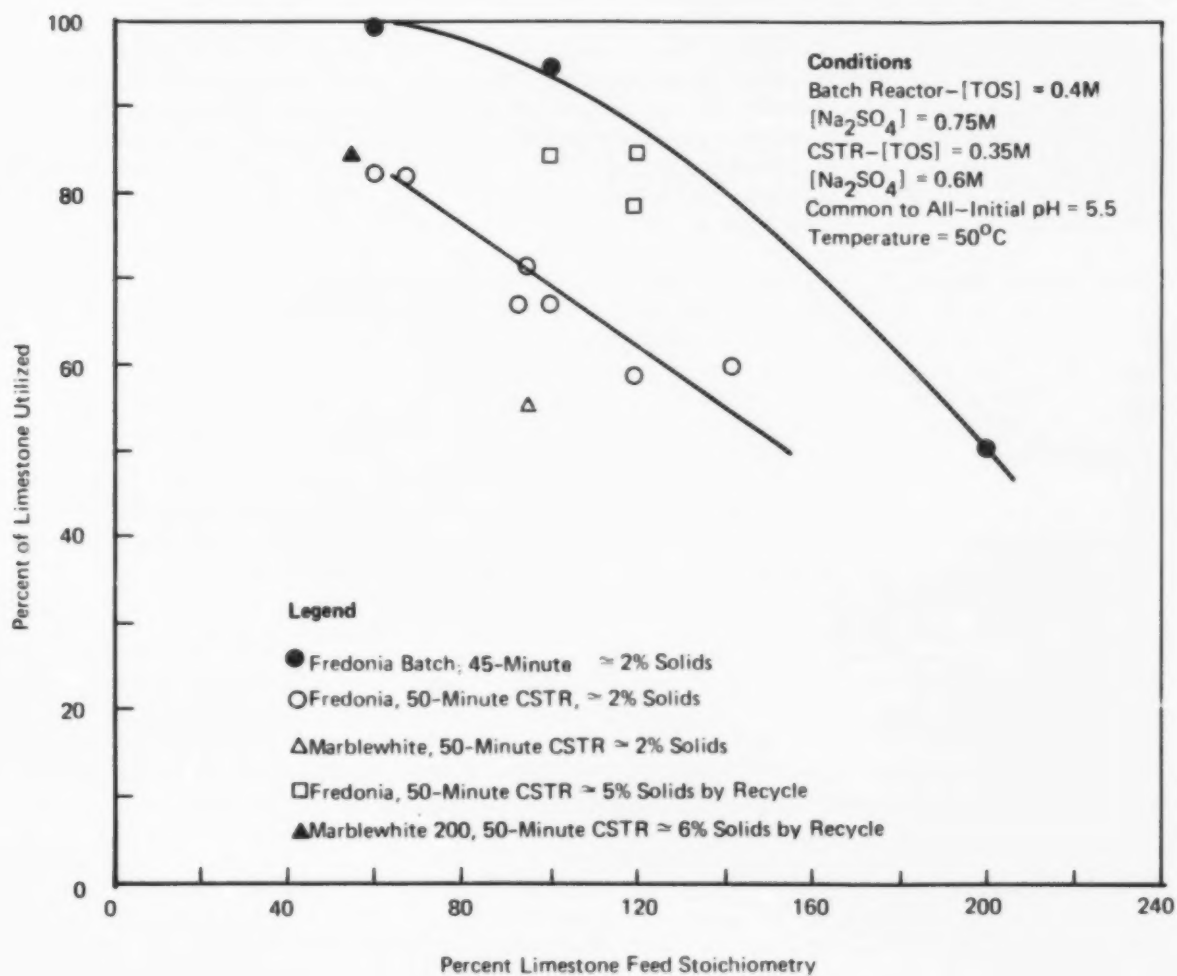


Figure V-5 Utilization of Two Limestones
in Batch and Continuous Reactors

The solids produced in the once-through CSTR experiments with Fredonia limestone were analyzed for their CaSO_4 content to ascertain whether or not sulfate was being precipitated during the course of the reaction. The amount of CaSO_4 found, expressed as a percent of the total CaSO_x present in the solids, is shown as a function of the Na_2SO_4 concentration in the reactor in Figure V-6. Significant amounts of sulfate precipitation were observed and, similar to the behavior observed when regenerating with lime, the sulfate precipitation increased linearly with increasing Na_2SO_4 concentration in the reactor solution. In fact, for a given Na_2SO_4 concentration, the actual sulfate precipitation with limestone was the same, within experimental error, as had been observed earlier for lime.

Because the matrix of CSTR experiments was limited in size, the sulfate precipitation shown in Figure V-6 was obtained from experiments with differing limestone feed stoichiometries as indicated in the figure. The fact that no significant deviations from linear dependence on sulfate concentration were observed for the wide variations in feed stoichiometry, suggests that stoichiometry had little or no effect on sulfate precipitation when limestone was used.

Total dissolved calcium levels measured in the reactor effluent for the experiments shown in Figure V-6 ranged from about 70 - 200 ppm, increasing with increasing Na_2SO_4 concentration. However, even at the highest sulfate levels studied, the solutions did not even closely approach saturation with respect to gypsum. By X-ray diffraction, the only sulfur-containing, crystalline phase which could be detected in numerous samples of product solids was $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$. These same two observations were made when regeneration with lime was studied and, when taken together with the essentially identical dependence of sulfate precipitation on Na_2SO_4 concentration, they indicate that sulfate precipitation by lime and limestone occurs by the same mechanism, possibly involving the formation of some sort of solid solution of CaSO_4 in the $\text{CaSO}_3 \cdot \frac{1}{2}\text{H}_2\text{O}$ lattice.

The solids produced in these initial CSTR experiments all exhibited settling behavior similar to that of "good settling material" previously produced in the laboratory and in the pilot plant when regenerating with lime.

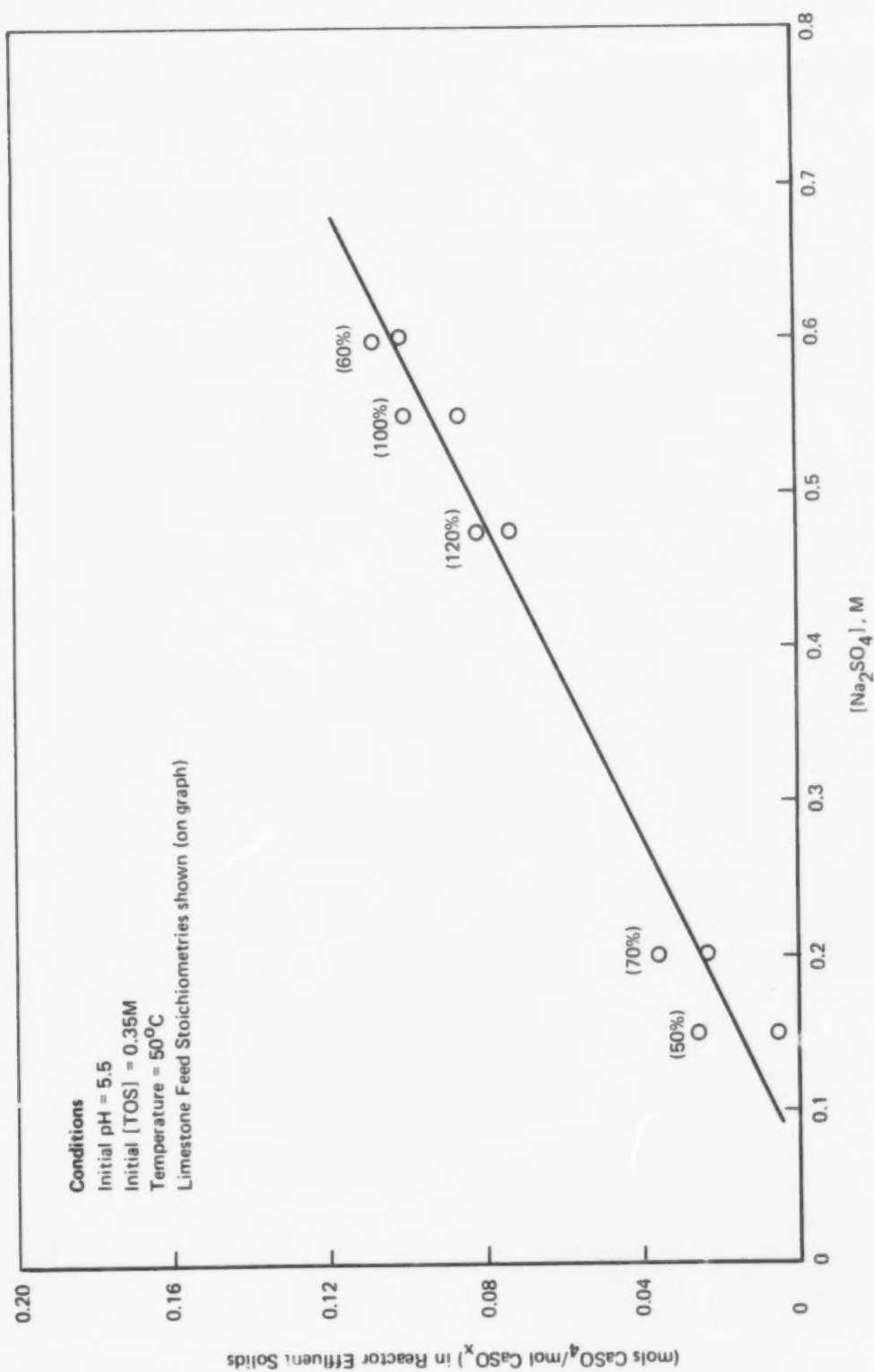


Figure V-6 Precipitation of Sulfate by Fredonia Limestone in 50-Minute CSTR Experiments

B. INITIAL PILOT PLANT TESTS AND SUBSEQUENT STUDIES OF FACTORS AFFECTING THE PHYSICAL PROPERTIES OF PRODUCT SOLIDS

With the laboratory experimental results up to that time all indicating that Fredonia limestone could be used to regenerate concentrated sodium bisulfite/sulfite solutions effectively, two closed-loop tests with Fredonia limestone were attempted in the ADL pilot plant. In the first run, the limestone reactor was a simple CSTR with a 90-minute holdup; in the second run the CSTR was operated with a recycle of solids from the thickener underflow. Although calcium utilizations ranged from 75% to 85%, the dewatering properties of the solids generated in the reactors were poor. Extremely low settling rates resulted in the carryover of a considerable amount of solids in the thickener overflow until, after a short period of operation, the scrubber feed liquor contained 2 - 3 wt % solids. The filter cake reflected the poor quality of solids. The washed cake averaged less than 40 wt % (dry cake basis).

During operating periods when the reactor effluent slurry pH fell below 6.8, bubbling began to occur in the thickener, indicating that a significant amount of additional reaction was taking place. This contributed to the solids carryover in the thickener overflow.

After this initial pilot plant experience, laboratory studies with the seemingly negative goal of producing poor settling solids were undertaken in an effort to explain the pilot plant results. Two obvious differences between operating conditions in the laboratory and in the pilot plant were explored initially in 50-minute CSTR experiments in which the solutions to be regenerated were similar to those used in earlier experiments (0.6M Na_2SO_4 and 0.36M TOS). Limestone had been fed as a slurry in the laboratory, but was fed dry in the pilot plant. A CSTR experiment was conducted in the laboratory with limestone fed dry and no deterioration of settling properties over those observed previously could be detected.

The sodium sulfite/bisulfite/sulfate solutions which were regenerated in the laboratory did not contain dissolved CO_2 , while in the pilot plant the scrubber effluent did contain dissolved CO_2 which could have possibly impeded the reaction. However, good solids settling behavior was still observed when the solution fed to the laboratory CSTR was saturated with CO_2 .

The one other significant difference between the laboratory and pilot plant experiments was the Na_2SO_4 concentration. During the pilot plant tests, the Na_2SO_4 level ranged from 0.9 to 0.8M. The higher Na_2SO_4 level fed to the laboratory CSTR up to that time was 0.75M which, after dilution by the limestone slurry, resulted in a maximum Na_2SO_4 concentration of 0.6M in the reactor. Subsequent laboratory CSTR experiments at higher concentrations of Na_2SO_4 indicated that the higher Na_2SO_4 level in the pilot plant was probably the factor responsible for the poor settling behavior observed.

The dramatic change in settling behavior which occurred when the concentration of Na_2SO_4 in the laboratory CSTR was changed from 0.6 to 1.0M is shown in Figure V-7. At intermediate concentrations of Na_2SO_4 (0.75 - 0.80M), the settling behavior observed for samples taken one hour after the reactor had filled and begun to overflow was intermediate between the two extremes in observed settling behavior. At those intermediate Na_2SO_4 levels, it was further observed that settling behavior tended to improve and approach that observed for 0.6M Na_2SO_4 as the reactor was allowed to continue to operate for an additional period of 2 -3 hours. However, the very poor settling behavior observed at the highest Na_2SO_4 concentration did not change over a period of several hours of reactor operation.

Clearly, on the basis of the 50-minute CSTR laboratory results to date, the need to control sodium sulfate concentrations at a level of 0.75M or lower appears to be an important prerequisite for successful system performance when regenerating with limestone. Whether this restriction can be relaxed if longer residence times are employed will be an important element of future CSTR tests.

One additional area related to the regeneration with limestone which is currently being studied is the effect of extraneous species on the regeneration reaction. Limestone can contain varying amounts of magnesium; chlorides can be absorbed from flue gases; and it is likely that smaller amounts of other soluble species could build up to significant levels during extended periods of closed-loop operation. To date, the effects of magnesium have been studied in batch and CSTR experiments; the effects of chloride and other species will be studied in future experiments.

The effect of magnesium, added as MgSO_4 at a concentration of 200 ppm (mg/liter) on the regeneration with Fredonia limestone in a batch experiment is shown in Figure V-8, where TOS removal as a function of time with and without the added magnesium is compared. The 200 ppm of Mg^{++} reduced the reaction rate significantly. In fact, if the calculated initial reaction rates observed with 200 ppm Mg^{++} present are compared with those shown in Figure V-3 when the Na_2SO_4 level was increased to 1.25M, it can be seen that the presence of the small amount of magnesium slowed the reaction more than increasing the Na_2SO_4 level from 0.75M to 1.25M.

The settling behavior of solids produced during 50-minute CSTR experiments in which about 100 and 200 ppm of Mg^{++} had been added to the solution to be regenerated is shown in Figure V-9. Settling curves for experiments in which no MgSO_4 was added are included for comparison. The 19- and 24-ppm Mg^{++} levels shown in Figure V-9 for these experiments were the amounts of magnesium dissolved from the limestone.

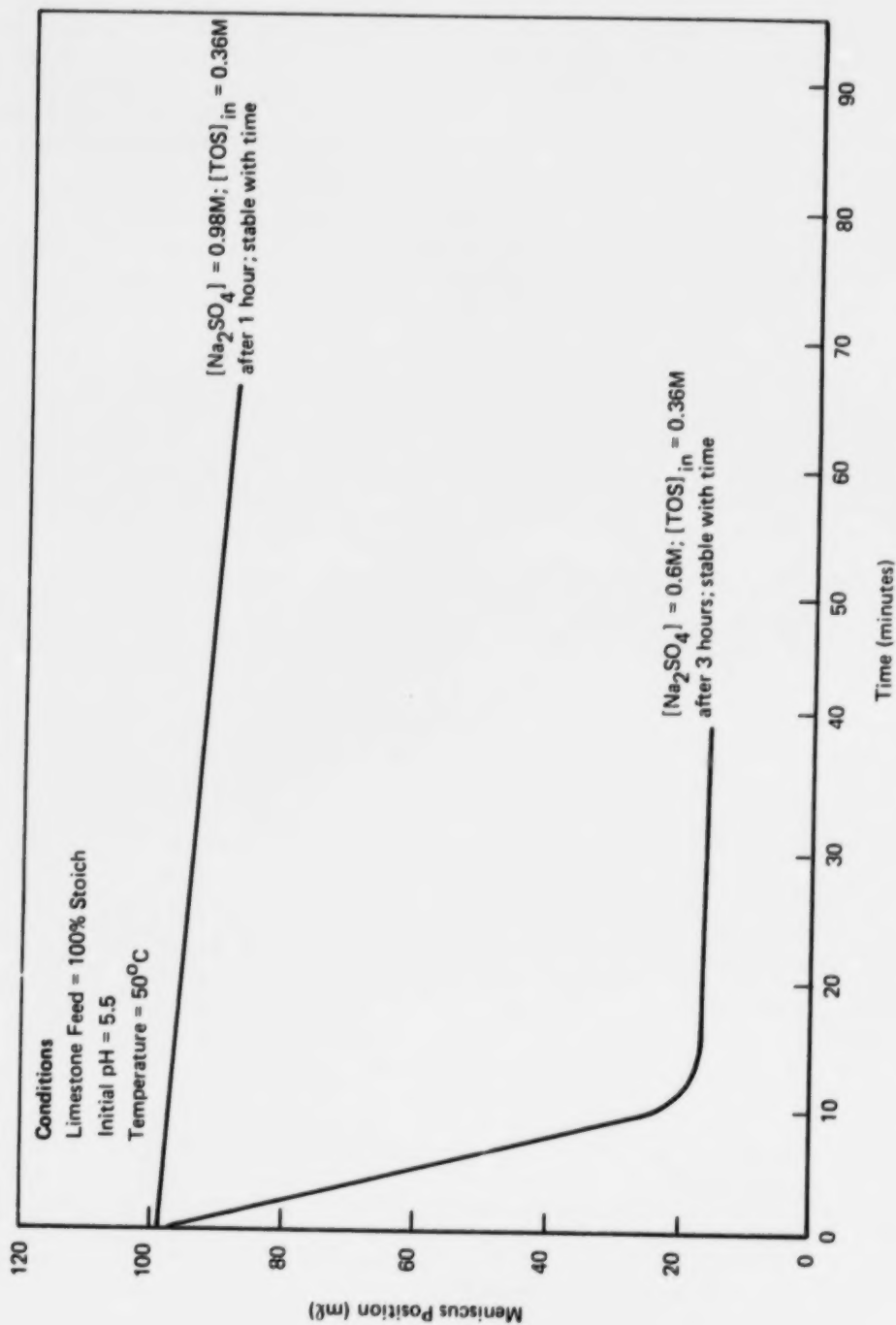


Figure V-7 Settling Curves For Solids Produced in 50-Minute CSTR Experiments Using Fredonia Limestone

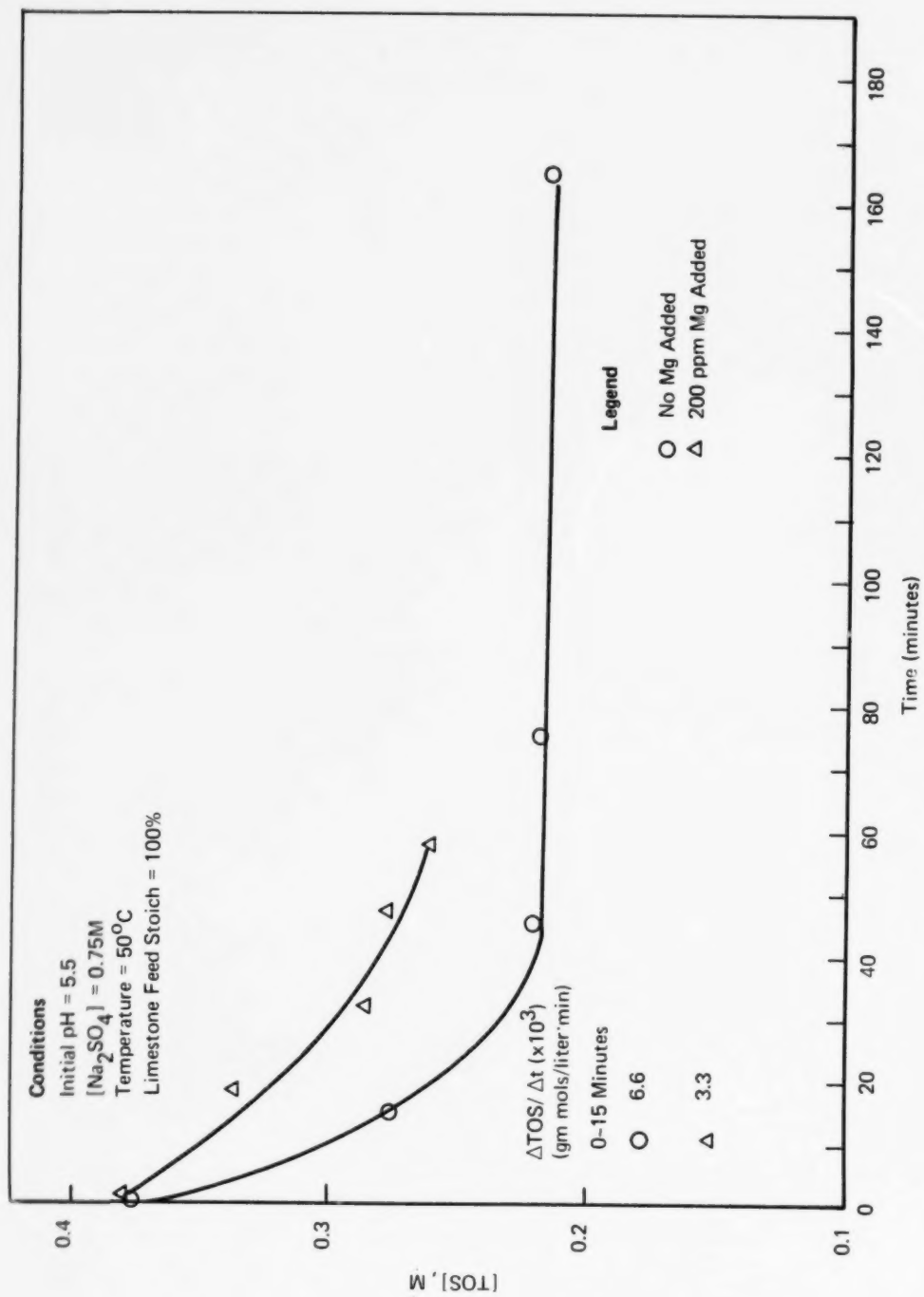


Figure V-8 Effect of Soluble Magnesium on Batch Regeneration Reaction Rates With Fredonia Limestone

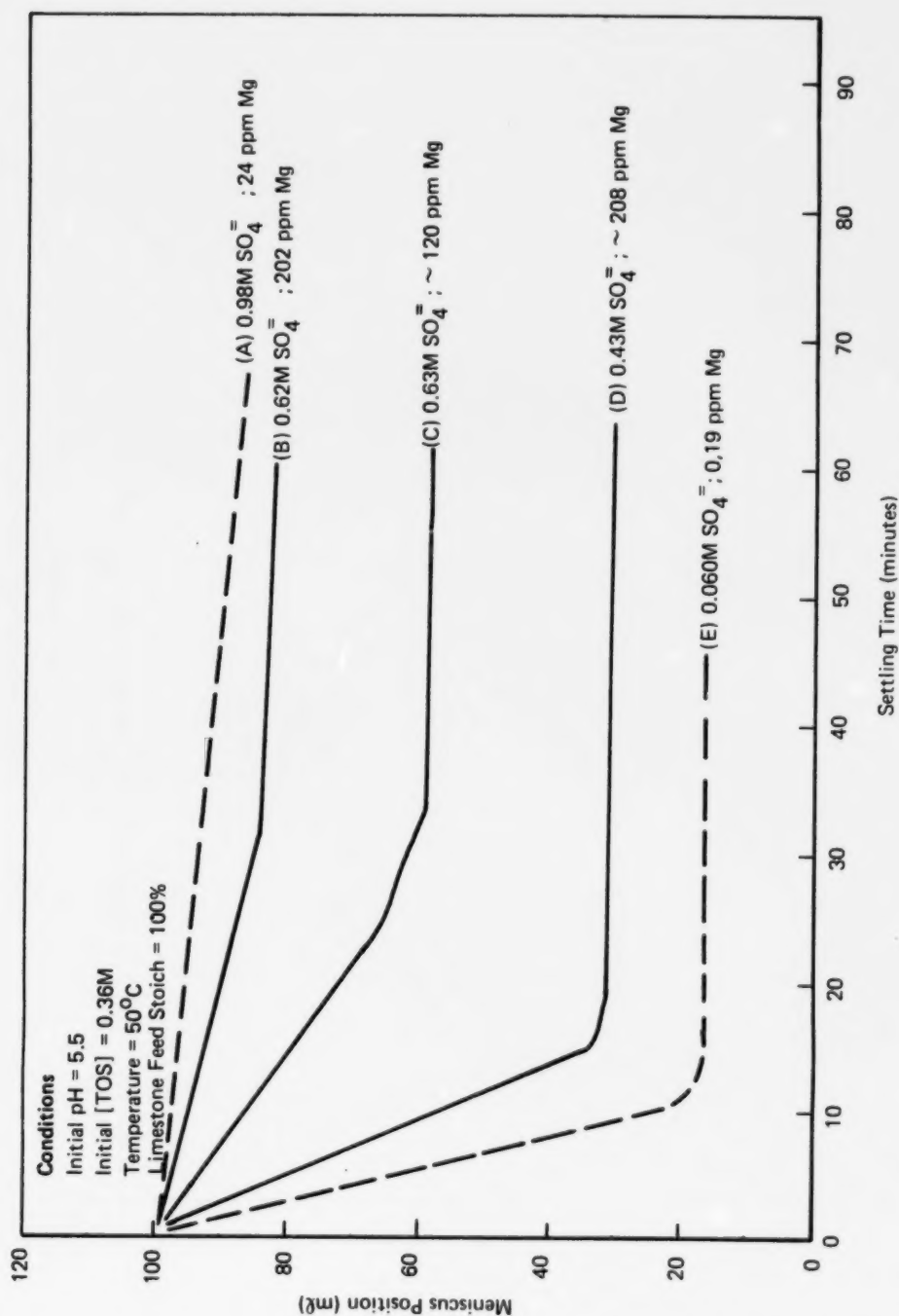


Figure V-9 Settling Behavior Observed For Solids Produced in a 50-Minute CSTR Using Fredonia Limestone

The significant change in settling behavior when Mg^{++} was added can be seen by comparing the settling curves (B, C and E) for the three experiments in which the concentration of Na_2SO_4 was held essentially constant at about 0.6M. The effects of Na_2SO_4 and Mg^{++} levels in solution appear to be similar and, after a fashion, additive. When the concentration of Na_2SO_4 was reduced to 0.43M (curve D), moderately good settling behavior was observed even with 208 ppm Mg^{++} present.

Additional CSTR studies of the regeneration reaction with limestone in the presence of magnesium at longer reactor residence times will determine whether or not the lower reaction rate can be accommodated given additional reaction time. If the additional studies confirm that soluble magnesium is really as deleterious as it appears to be at this time, and no easier means is found to circumvent or eliminate its effect, it should be possible to hold Mg^{++} at acceptable levels in the process loop by treating a portion of the process flow with lime to remove Mg^{++} as magnesium hydroxide. The results of initial batch experiments in which samples of the effluent slurry from the laboratory CSTR were treated with lime to various pH's are shown in Table V-1. At a final pH of 9.7, essentially no Mg^{++} was removed, but when the pH was raised to 11.2 a reduction in soluble Mg^{++} of about 98% was achieved. This approach may necessarily limit the process to the use of low magnesium limestones.

Table V-1

LIME TREATMENT^a OF
CSTR EFFLUENT SLURRY AFTER REGENERATION WITH LIMESTONE

----- Before Treatment -----			----- After Treatment -----		
<u>[TOS], M</u>	<u>[Mg], ppm</u>	<u>pH</u>	<u>[TOS], M</u>	<u>[Mg], ppm</u>	<u>pH</u>
.262	163	6.70	.188	162	9.7
.260	161	6.70	.190	3.4	11.2
.262	165	6.75	.179	2.7	12.05

^aBatch Reaction, 50-min reaction time; [Na₂SO₄], 0.61 M; Temp., 46°C.

C. CONCLUSIONS REGARDING LIMESTONE REGENERATION OF CONCENTRATED SODIUM BISULFITE/SULFITE SOLUTIONS

The results available to date from the ongoing laboratory program indicate that a sodium/calcium dual alkali process using limestone may be feasible. Using limestone in place of all or the majority of the lime which would otherwise be required should make the economics of sodium/calcium dual alkali FGD systems more attractive, if limestone can be accommodated within a reasonable degree of increase in process complexity or control requirements.

Because limestone is a less reactive regenerant than lime, and because it occurs naturally in a variety of forms which differ widely in their reactivity, a greater amount of overall attention will be required to ensure reliable, efficient process operation with limestone than with lime. The characteristics of regionally available limestones and the designs of processes to be sited in those areas will probably need to be considered together. The quality of incoming limestone will need to be assured to prevent process upsets, and considerably more attention may need to be given to the control of process liquor composition than would be necessary in a lime-based process.

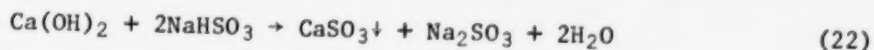
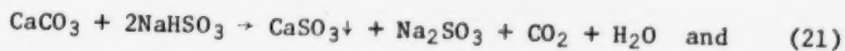
The investigation of the use of limestone in dual alkali processes is continuing and will proceed to the pilot plant if a viable approach is developed.

VI. LIMESTONE/LIME DILUTE MODES

As explained in Section I, above, operating with dilute sulfite/bisulfite (TOS) solutions represents one possibility for increasing the proportion of sulfate in the cakes produced during regeneration. Against this possible benefit must be balanced the cost penalties (represented by larger equipment and higher pumping rates) caused by operating with more dilute TOS solutions.

A. EPA LABORATORY RESULTS

In its Research Triangle Park laboratories, EPA had already performed some preliminary scouting runs on the regeneration of dilute TOS solutions with limestone and lime. These were summarized by Draemel⁹, who showed that the CaSO_3 exhibited a pronounced tendency to supersaturate; i.e., the reactions



did not proceed rapidly to completion in dilute TOS solutions. His scouting runs indicated that this tendency could be counteracted by recirculating the product solids through the reactor, and that maintaining a level of, say, 5% solids in the reactor might permit reasonable utilization (80-90% with lime, 50-70% with limestone), with 1-hour holdup times.

EPA extended this work by performing a series of continuous limestone runs in a 2-liter reactor followed by a 5-liter clarifier, with slurry recycle to the reactor. These were long-term runs (36-50 hours in length) at about 50°C. Satisfactory material balances were calculated for all, using an ADL computer program adapted for the purpose; this program performs balances around designated sections of the circuit for total weight, total volume, and the weight of water, sulfur, sulfate, sodium, and calcium.

The pertinent results for the 10 Fredonia limestone runs are summarized in Table VI-1. In all runs, a 17-20% CaSO_3 slurry was added to the TOS/ SO_4 feed solution, with a 1-hour residence time in the reactor and a 2-hour residence time in the thickener. Solids recycle was maintained from the thickener to the reactor to achieve the suspended solids concentrations shown in the table.

Table VI-1 EPA 2-LITER REACTOR CONTINUOUS FREDONIA LIMESTONE RUNS--SUMMARY

Run	Feed Solution Composition, M		Percent Solids in Reactor	Calcium Feed Stoichiometry	Calcium Utilization TOS Basis, a %	$\left(\frac{\text{SO}_4}{\text{CaSO}_4} \right)$ solids
	[TOS]	[SO ₄]				
21	0.14	0.25	2.1	1.13	90	10.1%
22	0.14	0.52	2.2	1.10	64	8.5
23	0.14	0.41	2.9	1.19	73	6.0
24	0.13	0.26	4.6	1.02	116	6.1
25	0.13	0.26	6.0	1.20	106	0.2
26	0.04	0.25	2.7	0.88	133	6.3
27	0.12	0.27	3.5	1.25	78	4.5
28	0.07	0.27	3.3	0.78	110	11.1
29	0.05	0.53	2.0	0.72	55	22.2
30	0.04	0.73	2.0	0.84	52	8.2

a. 100 x mols TOS Removed/mol Calcium Added

Note that attempts to increase the sulfate content of the cake by decreasing TOS and raising sulfate levels in solution met with only marginal success. Instead, this approach resulted in poor efficiency for TOS removal and poor Ca utilization. One interesting trend can be noted by observing the cake sulfate content in Runs 21, 24, 25, and 27, all conducted with essentially the same feeds and the same nominal stoichiometry, but differing in percent solids level in the reactor. There appears to be a downward trend in cake sulfate level as percent solids level (and thus effective solids residence time) in the reactor is increased.

On the other hand, lowering the TOS and raising the sulfate levels in solution did render a lime post-treatment more effective in precipitating sulfate from solution. In Runs 26 through 30, samples of the thickener overflow were agitated with excess lime for 1 hour, and the sulfate disappearance from solution was monitored. Table VI-2 shows the results for these runs, in terms of the ability of the system to reject sulfate as a fraction of total sulfur species precipitated, both for the lime step only and for the combination of the limestone and lime step (assuming all limestone step overflow to go on to lime treatment). The lime treatment solids in Runs 29 and 30 were analyzed and, as the table shows, the check with sulfate precipitation from solution is quite good.

The calcium and sulfate levels in the limestone reactor overflow and thickener underflow solutions of Runs 26 through 30 were multiplied for comparison with the apparent molal solubility product for $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ (predicted by the method of Kusik and Meissner⁵) for the ionic strength involved. Figure VI-1 shows that in all runs (Runs 21 through 30) levels were not high enough to precipitate $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ as such; however, whenever crystals of $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ were shaken with samples to test for calcium sulfate saturation, the product of the calcium and sulfate levels in the resulting solution rose to the predicted curve almost exactly (the squares on Figure VI-1). Figure VI-2 shows similar calculations for the calcium times the sulfite, compared with the analogous curve for $\text{CaSO}_3 \cdot 1/2\text{H}_2\text{O}$. In all runs, CaSO_3 supersaturation is indicated.

B. DILUTE MODE ALTERNATIVES

The results of this work have been utilized to help us project possible flow sheets for dilute operation at 35% sulfate regeneration rates. We have compared

- using limestone, with lime for added sulfate regeneration, and
- using lime, with sodium carbonate for added Ca softening.

The first of these has been called the "double loop-sulfite softening" mode because the calcium level leaving the lime reactor/clarifier is lowered by recycling the solution to the limestone reactor, in which the precipitation of CaSO_3 lowers the Ca level.

Table VI-2 EPA RUNS - POTENTIAL OF LIME POST-TREATMENT STEP

<u>LIME STEP ONLY</u>						
	Concentrations in in Feed Liquor, M		Concentrations in Treated Liquor, M		$\left[\frac{\Delta \text{SO}_4^{2-}}{\Delta (\text{SO}_4^{2-} + \text{TOS})} \right]_{\text{liquor}}$	$\left(\frac{\text{SO}_4}{\text{CaSO}_4} \right)_{\text{solids}}$
	$[\text{SO}_4^{2-}]$	$[\text{TOS}]$	$[\text{SO}_4^{2-}]$	$[\text{TOS}]$		
Run 26	0.2721	0.0184	0.2925	0.0090	(0)	-
27	0.2845	0.0648	0.2886	0.0180	(0)	-
28	0.2511	0.0418	0.2523	0.0214	(0)	-
29	0.5349	0.0326	0.5125	0.0100	49.8%	41.9%
30	0.7389	0.0313	0.6999	0.0106	65.3	64.5
<u>LIMESTONE PLUS LIME OVERALL</u>						
Run 26	0.2420 ^a	0.0390 ^a	0.2925	0.0090	(0)	-
27	0.2660	0.1183	0.2886	0.0180	(0)	-
28	0.2626	0.0712	0.2523	0.0214	17.1%	-
29	0.5208	0.0443	0.5125	0.0100	19.5	-
30	0.7183	0.0401	0.6999	0.0106	38.4	-

a. These (and following) adjusted for dilution of feed with limestone slurry.

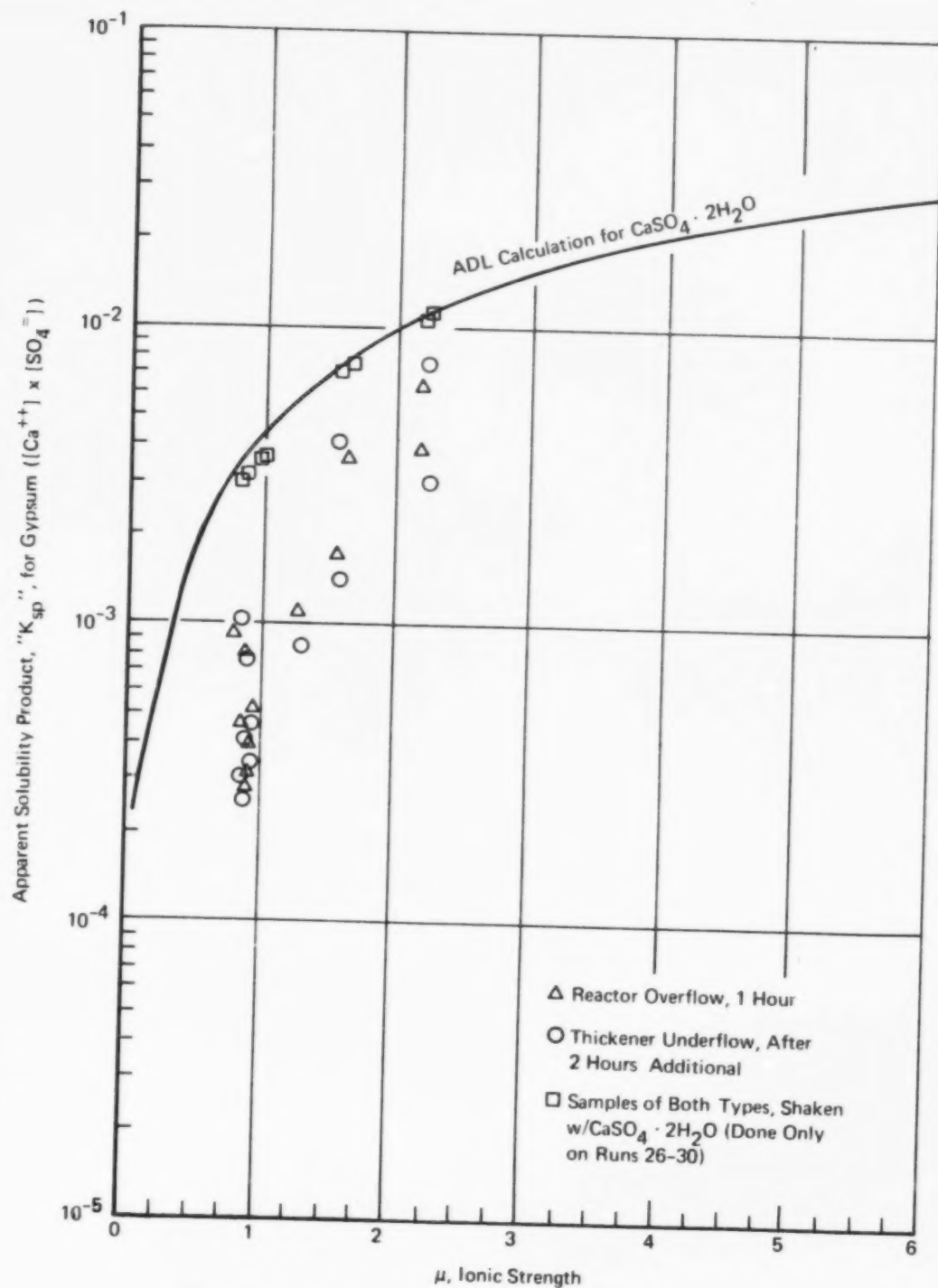


Figure VI-1 EPA Limestone Runs 21-30 Extent of $CaSO_4$ Under-Saturation

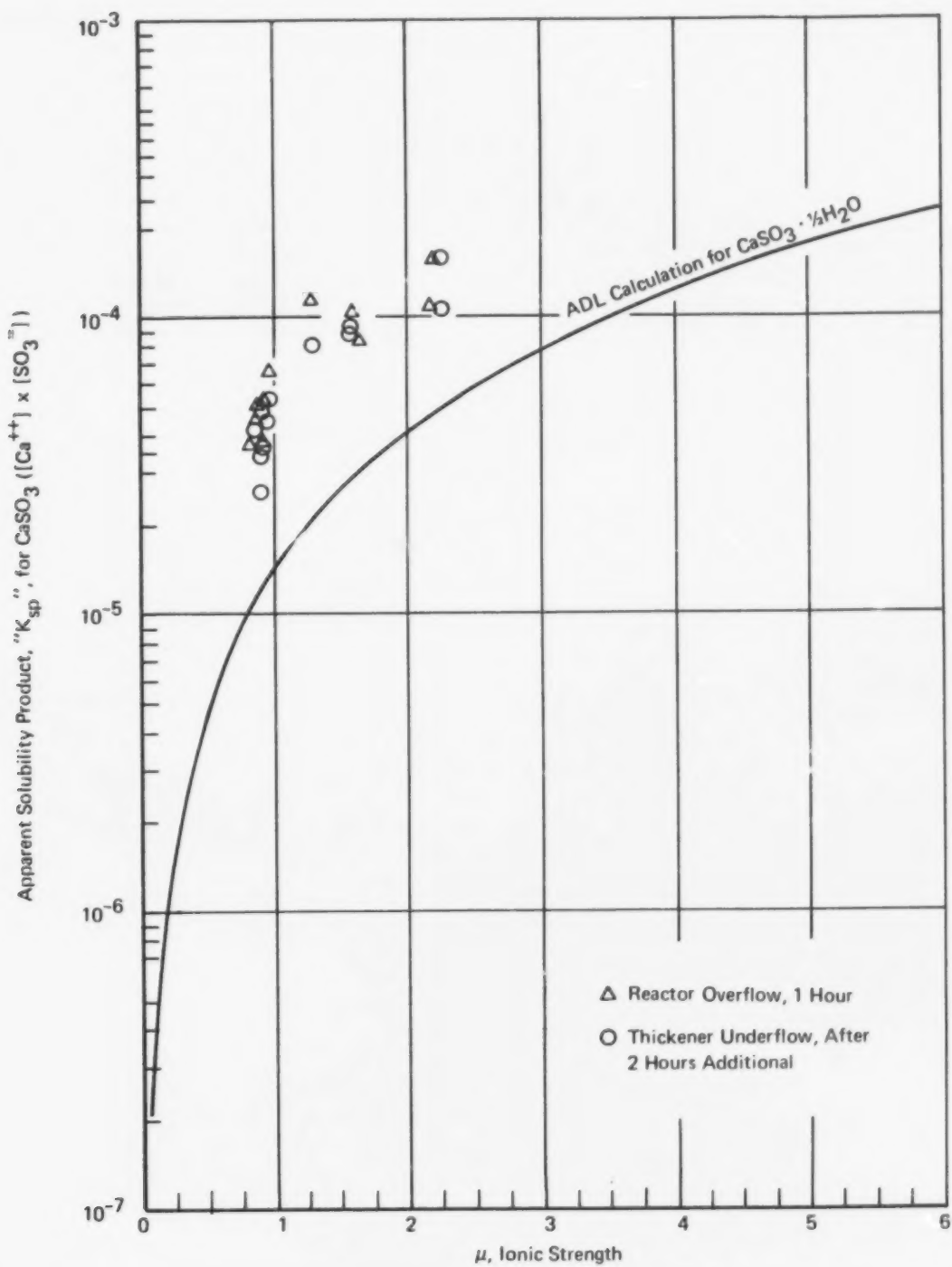


Figure VI-2 EPA Limestone Runs 21-30 Extent of CaSO₃ Supersaturation

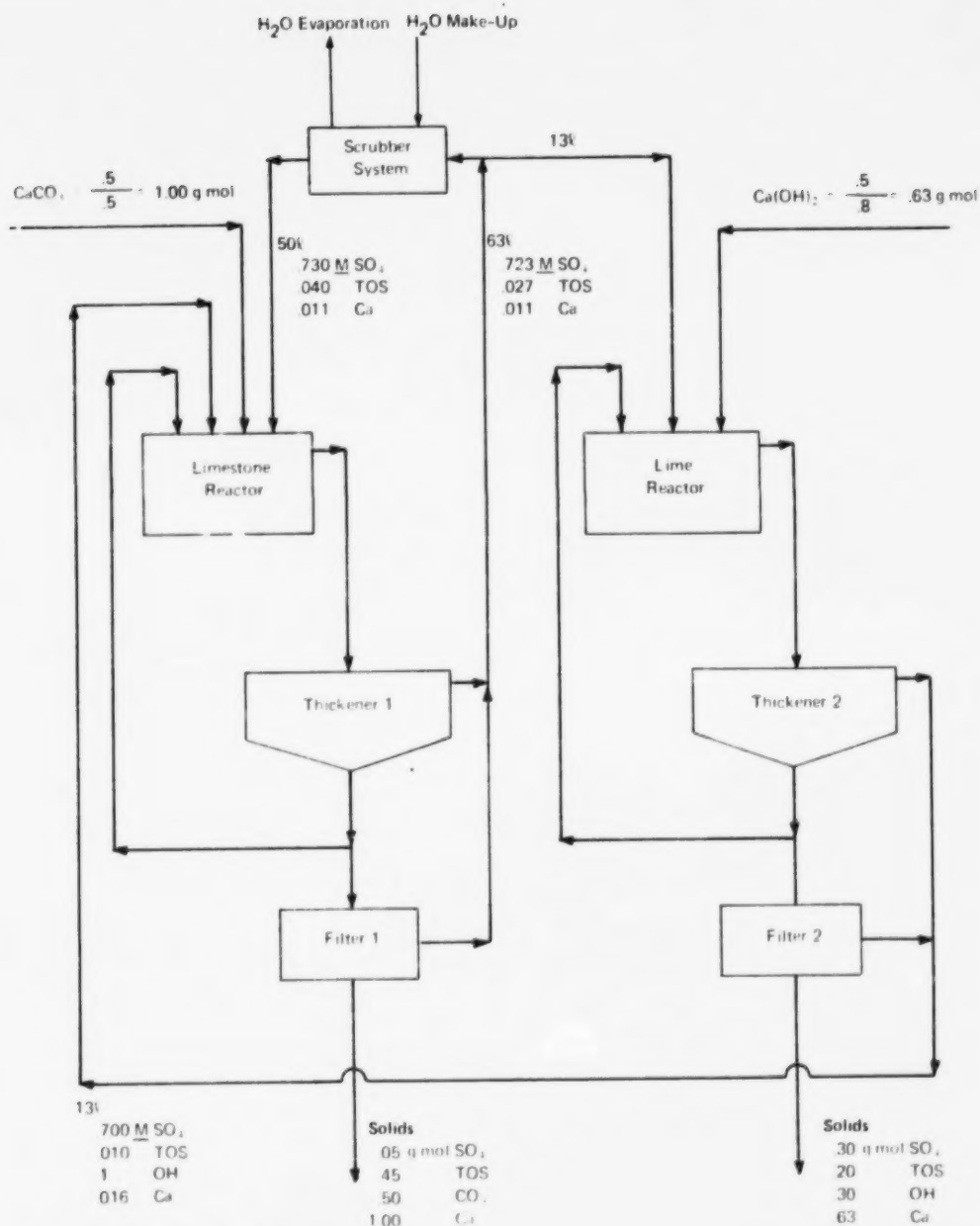
Figure VI-3 shows how one would operate a dilute limestone/lime system to precipitate sulfate to sustain a 35% oxidation level, based on the results of EPA Run 30 (confirmed by some preliminary ADL pilot plant runs not reported in this paper). About one-quarter volume of regenerated liquor would have to be sent to a lime loop (for sulfate precipitation) for every volume sent to the scrubber. At these conditions, the material returned to the scrubber would be undersaturated (with respect to $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$) by about 150 ppm Ca. The expected utilizations are only 50% for limestone and 80% for lime, based on experimental results. If one tries to keep up with higher levels of oxidation by sending higher fractions through the lime loop, the calcium undersaturation ("safety factor") decreases as the material leaving the lime reactor, which is saturated in $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$, begins to overpower the sulfite softening occurring in the limestone reactor; indeed, it does not appear possible to keep up with more than 40% oxidation without having to accept a regenerated scrubber liquor with a calcium level approaching within less than 100 ppm Ca of saturation with respect to gypsum.

Figure VI-4 shows how one might operate an alternative flow sheet using lime only, relying on sodium makeup with Na_2CO_3 for the necessary Ca softening. Again a 35% oxidation level has been taken as a basis. The expected $\text{Ca}(\text{OH})_2$ utilization is about 80%. The amount of sodium in the Na_2CO_3 needed for softening to a level 100 ppm below saturation is seen from Figure VI-4 to be less than 1% of the weight of the dry waste solids. Since soluble Na losses in well-washed cakes have been in the range of about 1%, the normal makeup of Na should introduce more than enough carbonate for softening.

The mode of operation in Figure VI-4 has a much smaller scrubber liquor feed; while the mode of Figure VI-3 requires 50 l/g mol SO_2 absorbed, that of Figure VI-4 requires 9.1 l/g mol SO_2 absorbed. The chief uncertainties in the mode of Figure VI-4 are:

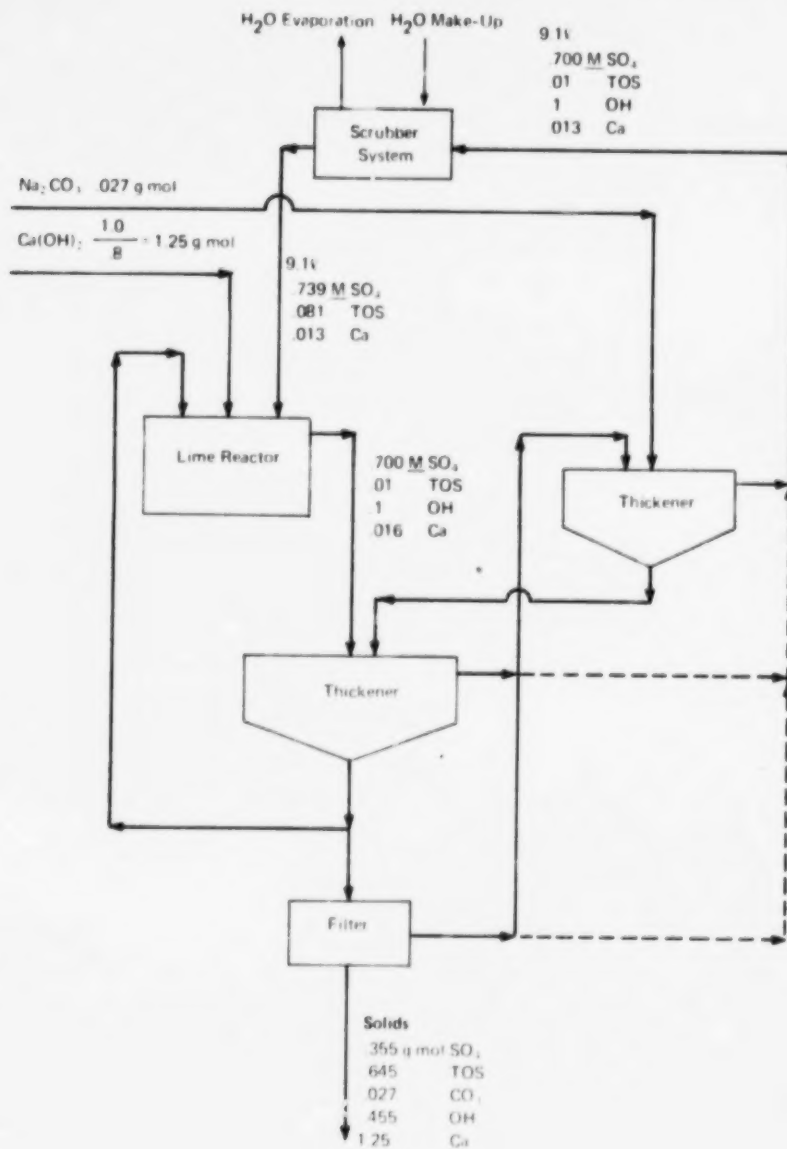
- 1) amount of solids recycle required to produce good settling and to eliminate CaSO_4 supersaturation in the lime reactor,
- 2) ability to produce good-settling carbonate solids in the carbonate treatment thickener, and
- 3) ability to bypass some liquor around the carbonate treatment step to save on the cost of the thickener (see dashed lines on Figure VI-4).

Based upon the schematic process flow sheets developed from the laboratory data, rough economic analyses were made comparing the two dilute systems: limestone/lime with sulfite softening and lime with soda ash softening. The limestone/lime system involves higher capital costs because of the higher flow rates through the absorption/regeneration system, but it has lower raw material costs since half of the regeneration is carried out



Basis: 1 g mol of SO_2 absorbed
 35% Oxidation Rate (all assumed in scrubber for convenience)

Figure VI-3 Dilute Limestone/Lime Schematic



Basis: 1 g mol. of SO₂ absorbed
 35% Oxidation Rate (all assumed in scrubber for convenience)

Figure VI-4 Dilute Lime/Carbonate Softening Schematic

using limestone, as opposed to complete regeneration using the more expensive lime in the alternate dilute system. It was assumed that both systems would have the same effective cake washing efficiency and require approximately the same sodium makeup. Because of lower yields on raw materials, the limestone/lime system generates more waste for disposal.

On the basis of the following analysis, it was found that the raw material cost savings for the limestone/lime system were not sufficient to offset the expected capital cost increases and increases in waste disposal costs.

Two cases were examined for the economic analysis:

- a 3.5% S coal application requiring 90% SO₂ removal, and
- a 1.0% S coal application requiring 80% SO₂ removal.

In both cases a relatively large cost difference was assumed between lime and limestone:

- 40 \$/ton, CaO
- 5 \$/ton, CaCO₃ (ground)

This large cost difference in materials favors the more capital-intensive sulfite softening approach. A waste disposal cost of

- 10 \$/ton waste (dry basis)

was used in all cases. A total annual fixed cost of 20% of the capital investment (including maintenance costs) was used to calculate annual costs.

Using this basis, we calculated the minimum capital investment difference at which the annual capital cost savings of the simpler lime approach equaled the annual operating cost savings of the limestone/lime approach. For the high-sulfur coal case, a capital investment savings of about \$7/kw or more would offset operating cost savings. In the low-sulfur coal situation, the corresponding figure dropped to only \$1.8/kw. In each case, we would expect the actual capital investments for the limestone/lime system to easily exceed those of the lime system by more than these calculated values. As a result, lime regeneration with soda ash softening was selected as the basic approach to be investigated for dilute modes in future pilot operations under this program.

VII. SOUTHERN SERVICES PROTOTYPE TEST PROGRAM

A. SYSTEM DESCRIPTION

Combustion Equipment Associates, Inc. (CEA) and ADL have jointly designed and built a dual alkali prototype system at Gulf Power Company's Scholz Station. The prototype system is designed to handle a flue gas rate equivalent to 20 Mw which will be drawn from one of the station's two 40 Mw coal-fired boilers.

The gas scrubbing system consists of a variable throat venturi in series with an absorption tower that can be operated either as a spray tower, tray tower (with up to four trays), or simply as a deentrainment separator. Ducting allows the flue gas to be taken through the newly installed electrostatic precipitator which will be operated at varying efficiency to change the dust loading to the scrubber system.

The absorber effluent treatment system includes a reactor system, a 40 foot thickener, and a rotary drum vacuum filter equipped with wash sprays. This equipment has been designed to operate at a full 20-Mw load with the boiler firing 5%S coal. The reactor system configuration also provides for the feeding of either hydrated lime or limestone, in either slurry and dry form.

A schematic showing the process equipment and configuration for this prototype system is provided in Figure II-2.

B. GENERAL OUTLINE

The test program will consist of two phases and will require approximately one year for completion. The test program will begin early in 1975 following system checkout and startup.

Phase I will last six months and will be directed toward collecting long term data to evaluate the overall operation of the process, including system operability and reliability. During this test phase, the Scholz Station boilers will be operated under "normal" plant conditions with the electrostatic precipitator in full service. Boiler loads will generally fluctuate from 50% to 100% of rated capacity and excess air rates will vary from about 20% to 45%. The sulfur content of the coal fired will range from 2 to 4% S and will vary according to normal coal shipment schedules and firing practices at the plant.

Throughout Phase I the dual alkali process will be operated in a concentrated active sodium mode using hydrated lime for absorbent regeneration. The process will run on a continuous around-the-clock basis with planned shut-downs coinciding only with boiler down time.

Phase II is designed to obtain shorter term data for a number of different system operating conditions. For planning purposes, this phase has been broken down into four test periods, each lasting from four to six weeks. The exact variables and system parameters to be examined in each of these periods have not as yet been fully defined; however, it is expected that the major variables considered in Phase II would include the use of different coals, the effects of fly ash, and possibly the use of limestone as the absorbent regeneration reactant. Enough flexibility will be maintained in this second phase to allow for the modification of test schedules and test conditions as particular needs arise. As in Phase I, the system will be operated in the concentrated active sodium mode.

Each of the test phases will involve sufficient sampling and data logging to support detailed material balances around the entire system and around specific process sections as desired. The main focus of the data collection and analysis will be the characterization of the process operation in terms of:

- scrubber efficiency
- oxidation rates
- sulfate regeneration
- solids properties
- sodium losses and
- lime and other chemicals utilization.

GLOSSARY

Active Sodium - Sodium associated with anions involved in SO_2 absorption reactions and includes sulfite, bisulfite, hydroxide and carbonate/bicarbonate. Total active sodium concentration is calculated as follows:

$$[\text{Na}^+]_{\text{active}} = 2 \times ([\text{Na}_2\text{SO}_3] + [\text{Na}_2\text{CO}_3]) + [\text{NaHSO}_3] + [\text{NaOH}] + [\text{NaHCO}_3]$$

Active Sodium Capacity - The equivalent amount of SO which can be theoretically absorbed by the active sodium. Active sodium capacity is defined by:

$$[\text{Na}^+]_{\text{active capacity}} = [\text{Na}_2\text{SO}_3] + 2 \times [\text{Na}_2\text{CO}_3] + [\text{NaOH}] + [\text{NaHCO}_3]$$

Calcium Utilization - The percentage of the calcium in the lime or limestone which is present in the solid product as a calcium-sulfur salt. Calcium utilization is defined as:

$$\text{Calcium Utilization} = \frac{\text{mols } (\text{CaSO}_3 + \text{CaSO}_4) \text{ generated}}{\text{mol Ca fed}} \times 100\%$$

Concentrated Dual Alkali Modes - Modes of operation of the dual alkali process in which the active sodium concentrations are greater than 0.15M active sodium.

CSTR--Continuous Feed Stirred Tank Reactor - A well agitated, baffled reactor vessel having a uniform concentration of species throughout. At any time the concentrations in the effluent from a CSTR are equivalent to those within the vessel.

Dilute Dual Alkali Modes - Modes of operation of the dual alkali process in which the active sodium concentration is less than or equal to 0.15M active sodium.

Sulfate Formation - The oxidation of sulfite. The level of sulfate formation relative to SO_2 absorption is given by:

$$\text{Sulfate Formation} = \frac{\text{mols } \text{SO}_3^- \text{ oxidized}}{\text{mol } \text{SO}_2 \text{ removed}} \times 100\%$$

Sulfate Precipitation - The formation of $\text{CaSO}_4 \cdot \text{XH}_2\text{O}$ in soluble solids. The level of sulfate precipitation in the overall scheme is given by the ratio of calcium sulfate to the total calcium-sulfur salts produced:

$$\text{Sulfate Precipitation} = \frac{\text{mols } \text{CaSO}_4}{\text{mol } \text{CaSO}_x}$$

TOS--Total Oxidizable Sulfur - Equivalent to the sum of all sulfite and bisulfite species.

REFERENCES

- ¹Draemel, D.C. and Kaplan, N., An Overview of Sodium-Based Double Alkali Processes, Proceedings of Flue Gas Desulfurization Symposium--1973, EPA-650/2-73-038, (December 1973).
- ²LaMantia, C.R., et al, Sulfur Dioxide Control Process Study--Sodium Scrubbing with Lime Regeneration, Report to the Illinois Institute for Environmental Quality, (January 1972).
- ³LaMantia, C.R., et al, Dual Alkali Process for Sulfur Dioxide Removal, CEP, 70, 66 (June 1974).
- ⁴Borgwardt, R.H., Preliminary Report of Test Results from EPA Alkali Scrubbing Test Facilities at the TVA Shawnee Power Plant and at Research Triangle Park, EPA Briefing, Research Triangle Park (December 1973).
- ⁵Meissner, H.P. and Kusik, C.L., I&EC Proc. Des. Div., 12, 205 (1973).
- ⁶Marshall, W.L. and Jones, E.V., J. Phys. Chem., 70(12), 4028 (1966).
- ⁷Kaplan, N., An Overview of Dual Alkali Processes for Flue Gas Desulfurization, EPA Flue Gas Desulfurization Symposium, Atlanta, (November 1974).
- ⁸Drehmel, D.C., Wet Limestone Scrubbing of Sulfur Oxides; Limestone Selection, Proceedings of the Sixty-Fifth Annual APCA Meeting, June 18-22, 1972.
- ⁹Draemel, D., Regeneration Chemistry of Sodium-Based Double-Alkali Scrubbing Processes, EPA-R2-73-186, Control Systems Division, Research Triangle Park, N.C. (March 1973).

APPLICABLE CONVERSION FACTORS
ENGLISH TO METRIC UNITS

VOLUME

$$1 \text{ gallon} = 3.785 \text{ liters} = 0.003785 \text{ m}^3$$

$$1 \text{ ft}^3 = 28.32 \text{ liters} = 0.02832 \text{ m}^3$$

DISTANCE

$$1 \text{ ft} = 0.3048 \text{ meters}$$

TEMPERATURE

$$^{\circ}\text{C} = (^{\circ}\text{F} - 32) \times 5/9$$

$$1^{\circ}\text{F} = 5/9^{\circ}\text{C}$$

ENVIRONMENTAL PROTECTION AGENCY
Technical Publications Branch
Office of Administration
Research Triangle Park, N.C. 27711
OFFICIAL BUSINESS
AN EQUAL OPPORTUNITY EMPLOYER

POSTAGE AND FEES PAID
ENVIRONMENTAL PROTECTION AGENCY
EPA - 335
SPECIAL FOURTH-CLASS RATE
BOOK



Return this sheet if you do NOT wish to receive this material ☐,
or if change of address is needed ☐. (Indicate change, including
ZIP code.)

PUBLICATION NO. EPA-650/2-74-126-a

B202
H